Energy, Mines and Resources Canada
Énergie, Mines et Ressources Canada

CANMET
Canada Centre for Mineral and Energy Technology
Centre canadien de la technologie des minéraux et de l'énergie

PROCEEDINGS OF THE MEETING OF THE CANADIAN URANIUM PRODUCERS' METALLURGICAL COMMITTEE HELD IN BANCROFT, MAY 19-20, 1982

G.M. RITCEY
Extractive Metallurgy Laboratory

MARCH 1983

Project - 4.4.0.2.01 - Improvements to Conventional Uranium Extraction Technology

236 pp

ENERGY RESEARCH PROGRAM
MINERAL SCIENCES LABORATORIES
DIVISION REPORT ERP/MSL 83-39(TR)

PREFACE

This publication has been prepared to provide a permanent record of the Proceedings of the 1982 Annual Meeting of the Canadian Uranium Producers' Metallurgical Committee which was held in Bancroft, Ontario. In addition to the Canadian attendees, the meeting was fortunate in having members of the IAEA attending. This group had a meeting earlier the same week in Ottawa on "Migration of Radium and Other Contaminants Present in Liquid and Solid Wastes from Mining and Milling of Uranium". The author's original unedited technical presentations together with the discussions have been incorporated into this report.

The next annual meeting of the Committee is scheduled for May 1983 and will be held in Toronto, Ontario.

ACKNOWLEDGEMENTS

The participation of those who attended is to be commended and is the main factor that demonstrates the usefulness of the meeting. The generous hospitality provided by Madawaska Mines certainly added to the success of the meeting. Acknowledgement is also made to Mrs. L. Paquette, CANMET, who assisted in the preparation of this record of the meeting.

G.M. Ritcey
Secretary, CUPMC

BUSINESS SESSION

A short business meeting concluded the two day meeting. The bank balance was stated as $242.46 as of April 30, 1982.

The next meeting of the CUPMC was set for May 1983 in Toronto, Ontario, with Ontario Research Foundation the host, and Lacky Lakshmanan as Chairman.

Madawaska Mines was thanked for providing the arrangements and hospitality of the meeting and the tour of the Madawaska Mill.

CANADIAN URANIUM PRODUCERS' METALLURGICAL COMMITTEE
1982 MEETING, BANCROFT, ONTARIO

PLACE: YE OLDE BARN RESTAURANT (HIGHWAY NO. 62N, 1 MILE NORTH OF BANCROFT)

DATES: MAY 19 AND 20, 1982

CHAIRMAN: LEO MCFARLANE, MADAWASKA MINES

WEDNESDAY, MAY 19,1982

0930 - 1000 a.m.	Coffee (Courtesy Madawaska Mines Ltd.)
1000	Milling high grade uranium ore at Cluff Lake - P. Fournier,Amok
1040	"Caro's Acid" - A New Oxidant for Acid Leaching of Uranium Ores - R. Ring, Australian Atomic Energy Comm.
1110	Midwest Project Update - Metallurgy and Waste Treatment - L. Melis, Canada Wide Mines
1200 - 1330	Lunch at Ye Olde Barn Restaurant
1330	Status Report on Kidd Creek Aquitaine Project in Nova Scotia, R. John, Kidd Creek Mines
1410	R&D Activities at Eldorado Nuclear Limited - An Overview - I. Itzkovitch, Eldorado Nuclear Limited
1440	Report of IAEA and Fort Collins Meetings - M.C. Campbell, CANMET
1510 - 1530	Coffee (Courtesy Madawaska Mines Ltd.)
1530	Uranium and Rare Earths in China - G.M. Ritcey, CANMET
1800	Dinner at Ye Olde Barn Restaurant (Courtesy of Madawaska Mines Ltd.)

THURSDAY, MAY 20, 1982

0900	Evaluation of Tailings Stabalization Methods - V. Lakshmanan, ORF
0940	Composting of Municipal Wastes for Tailings - CANMET staff
1010	Business Meeting
1030	Coffee
AM or PM	Tour of Madawaska Mill and Tailings

CUPMC ATTENDANCE, MAY 19-20, 1982

NAME	AFFILIATION
L. Bates	Kilborn Engineering
P. Benes	Technical University of Prague, Czechoslovakia
W. Bonisteel	Simon Carves
H. Branstatter	Ontario Research Foundation
M. Campbell	CANMET
G. Cleland	Witteck Development
W. Craigen	CANMET
M. Cristovici	CANMET
A. D'Agostino	Eldorado Nuclear, Port Hope
B. Donati	C.N.E.A., Argentina
T. Edwards	Simon Carves
J. Evans	A.A.E.C., Australia
G. Feasby	Eldorado Nuclear, Beaverlodge
N. Fourcade	C.N.E., France
P. Fournier	Amok
J. Fox	Wright Engineers
A. Gallant	Denison Mines
G. Gee	U.S. N.R.C., USA
J. Goode	Kilborn Engineering
W. Gow	CANMET
V. Haw	CANMET
A. Hayden	A.H. Ross & Associates
K. Hester	Rio Algom Mines
L. Heymann	Agnew Lake Mines
I. Itzkovitch	Eldorado Nuclear, Ottawa
R. John	Kidd Creek Mines
M. Kalin	University of Toronto
Y. Kitahara	Power Reactor & Nuclear Fuel Corp., Japan
V. Lakshmanan	Ontario Research Foundation
R. Laliberte	Denison Mines
C. Lendrum	Consulting Engineer
M. Lopez	Interox America, USA
L. MacFarlane	Madawaska Mines
L. Marple	Texas Dept. of Health, USA
I. Masters	Sherritt-Gordon
A. Mattus	Interox America, Texas
L. Melis	Canada Wide Mines
W. Milde	Gulf Minerals
J. Molinari	IAEA, Austria
R. Oberholtzer	Eldorado Nuclear, Port Hope
A. Paschoa	Pontificia Universidade Catolica, Brazil
M. Raghavayya	Uranium Corp. of India, India
R. Ring	A.A.E.C., Australia
G. Ritcey	CANMET
D. Rollwagen	Madawaska Mines
B. Rosner	Key Lake Mining
J. Schisler	Eldorado Nuclear, Port Hope
R. Williamson	Lakefield Research
M. Yetman	Rio Algom Mines
P. Zettwoog	C.N.E., France

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- R.J. Ring, D.J. Waters and M. Yamine --------------------- 2-1 - 2-20
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The Midwest Uranium Project - Development of the Milling
Process - L.A. Melis, K.S. Fraser and V.I. Lakshmanan ------ 3-1 - 3-78
- Discussion --- 3-79

Status Report on Kidd Creek Aquitaine Project in
Nova Scotia - R. John -------------------------------------- 4-1 - 4-9
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R&D Activities at Eldorado Nuclear Limited - An Overview
- I. Itzkovitch -- 5-1
(No paper available)

Report of IAEA and Fort Collins Meetings - M.C. Campbell ----- 6-1 - 6-7

Uranium and Rare Earths in China - G.M. Ritcey --------------- 7-1 - 7-16
- Discussion --- 7-17
(no notes taken)

Evaluation of Tailings Stabilization Methods - G.M. Ritcey,
J.M. Skeaff, V.I. Lakshmanan and L.M. Luckevich ------------ 8-1 - 8-30
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Compost Production from Municipal Wastes of Canadian
Mining Towns - A. Jongejan --------------------------------- 9-1 - 9-24
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MILLING HIGH GRADE URANIUM ORE AT CLUFF LAKE, SASKATCHEWAN

by Ph. Fournier, Project Manager at Amok/Cluff Mining

0. INTRODUCTION

The "D" orebody was discovered in the Cluff Lake area in 1970. Extensive drilling was done in the following years to assess the quantity and quality of this unique deposit and also to provide samples for laboratory metallurgical testing. In 1974, it was recognized that about 5,000 Tonnes of uranium were there, in a close to the surface deposit with an average grade of 7% in uranium metal (equivalent to 160 pounds of U_3O_8 per ton).

Facing an unusual problem, not conventional processes were tested including a nitric acid leaching in order to separate the radium from the gangue which proved not too successful. Also core samples showed that extreme variations of grade would be encountered, from a low 1% to a high 45%. This meant that either a highly sophisticated blending method would be required or a dual process would have to be found. Fortunately, the uranium mineralization was entirely made of pitchblend and uraninite with high specific gravity which allowed to upgrade by gravimetric method the poorer fraction up to a grade compatible with the richer fraction.

This process was basically selected in early 1976 but the project was shelved for 2 years pending the result of a uranium inquiry in the province of Saskatchewan. The "Bayda Commission" released its final report in spring 1978 with favorable recommendations for the project to proceed conditionally to stringent environmental and social conditions. These recommendations were taken up by the provincial and federal regulatory bodies so that a surface lease agreement was signed in October 1978 allowing the project to be launched.

Construction took place between March 1979 and October 1980 and actual startup with high grade ore was done in late January 1981. The normal capacity was reached in May 1981 and the plant was able to produce 3.2 million pounds of U_3O_8 in 1981. On the other hand, the D deposit was completely mined out in two summer seasons and the ore was then stockpiled.

1. PROCESS DESCRIPTION

1.1 Grade

As said before the average grade of 7% uranium does not reflect the large fluctuations of the grades which the mill has to deal with. Following technical and economical considerations, the following cutoffs have been selected:-

- Below 0.03% is considered as waste.
- The fraction 0.03% to 0.3% is stored on a separate stockpile, for processing at a later date.
- The fraction 0.3% - 20% is the so-called "gravimetric grade" meaning that it is processed in the gravimetric separation circuit. This fraction corresponds to more than 54,000 Tonnes at about 4%. For storage considerations this fraction has been separated in two piles grading respectively 1.8% and 16%.
- The fraction above 20% is the so-called "direct leach grade" meaning that it bypasses the gravimetric separation circuit and goes directly to the grinding and leaching. This fraction represents some 9,000 tonnes of ore at about 29%.

Typical analysis of the two types of ore are given in Table 1.

1.2 Ore receiving, crushing and storage (ref. figure 1)

The ore is delivered by 15 ton covered trucks, dumped into a 30 ton truck dump hopper and belt conveyed to a primary 24 x 36 jaw crusher. The discharge of the jaw crusher at minus 4 inches is belt conveyed to one of the 3 ore storage buildings according to the grade. The capacity of these storages is respectively 2,000 Tonnes for the direct leach grade ore and 10,000 Tonnes for the other two.

The first one is made of concrete to provide shielding against the gamma radiation. The other two are made of backfill and are covered. From the bins, the ore is withdrawn by vibrating feeders and conveyed to the secondary crushing circuit that operates in a closed circuit on a Tyler vibrating screen. The crusher is a 4'1/4' shorthead Symons cone crusher. The fine ore is then conveyed to 2 x 200 Tonnes concrete made fine ore bins.

A typical screen analysis of the crushed ore feeding the gravimetric separation is given in table 2.

The capacity of this section is severalfold what is required for the leaching plant. This allows the primary crushing and the ore haulage to operate essentially in summer and the secondary crushing to operate on a demand basis on day shift, only a few hours per week for the direct leach ore and a few hours per day for the gravimetric ore according to the type of operation.

1.3 Gravimetric Separation (ref. figure 2)

This section started in July 1981 and ran for 5 months before being shutdown for winter. Some modifications were made during the shutdown period to increase both the capacity and the recovery. Operations have resumed in mid April 1982.

Basically the gravimetric separation involves jigging and tabling the ore according to its size to separate the heavy pitchblend from the lighter material, mainly silico aluminate. Figure 3 shows some typical results of this operation with various feed grades and in summary we can see that the gravimetric separation yields concentrates grading above 30% uranium while the tails are about 0.8%. Recoveries of 80 to 85% are reached.

From the fine ore bins, the ore is first screened at 6 mesh, the plus fraction being sent to a Yuba jig while the minus 6 mesh is sent to a Denver mineral jig. To avoid carry over of very fine clayish particles the feed to the Denver jig is cycloned to remove the minus 200 mesh fraction.

Both jigs give concentrates and tails. The concentrates report to the grinding circuit while the tails are again screened at 20 mesh. The plus 20 mesh fraction is stored for processing at a later date while the minus 20 mesh fraction is divided by a high frequency Derrick screen to feed 2 sets of concentrating tables, one set for the sands (-20, +100 mesh) one set for the fines (-100 mesh). Each set has two tables, the second one processing the middlings from the first. We have then a total of 4 tables.

The concentrates from the tables report to the grinding circuit, while the tails are thickened and stored for processing at a later date along with low grade ores from other orebodies.

1.4 Grinding

At this point there is no longer any distinction between the direct leach ore and the gravimetric ore and the grinding circuit is fed with either direct leach ore or gravimetric concentrates.

At this point also the capacity of the plant is 1.2 Tonne/hour with a grade around 25 to 30%.

The grinding circuit consists of a small hopper, a constant volume feeder, a 3'0 x 5'L Hardinge rod mill operating in close circuit with a 12 inch spiral classifier. The system grinds the ore (or concentrate) to about 350 microns that constitutes the feed to the leaching. The overflow from the classifier is pumped to a concrete made concentrate thickener and a concentrate slurry holding tank where a 65% solid slurry is stored.

1.5 Leaching and liquid solid separation (ref. figure 4)

Leaching is done in a series of 7 mechanically agitated tanks 4 feet in diameter and 6 feet high encased in 14 inches thick concrete for shielding against the radiation. The vessels and the agitators are rubber lined to withstand strong corrosive conditions.

A premixing tank is provided for sodium chlorate addition and the total retention time is 3.5 hours. Sulfuric acid is added in the first and the third leaching tanks controlled by conductivity.

There are several unique features in this leach system

- Sulfuric acid consumption is very high from a low 250 kg to a high 400 kg/Tonne of ore according to the grade. (see Fig 5)

- Sodium chlorate addition is very high also averaging 90 kg/Tonne. However the consumption of those two reagents is low per kilogram of uranium dissolved during the leach (1.5 and 0.4 kg/kg respectively).
- Although the leaching temperature is 55°C, there is no need for heat in the system. Actually we have to cool the first leach tank to maintain the desired temperature! (this is due to the sulfuric acid heat of dilution).
- While starting at 55% solids, the slurry density decreases to 20 to 25% solids at the end of leaching because of the amount of dissolved material.
- A very high concentration of dissolved uranium is obtained usually above 100 gpl.

Liquid solid separation is done by filtration using two 7.5m^2 horizontal Philippe belt filters, one in operation and the other in standby although they can also be operated in series with a repulping tank in between.

Washing is done on the filters with a high washing ratio and the cake is discharged at about 35% moisture.

Leaching and washing efficiencies are excellent with figures above 99.7% for leaching and 99.5% for washing easily obtained. The leaching and filtration provides about 25 GPM of leach filtrate at about 40 to 45 gpl uranium to feed the precipitation steps.

1.6 Iron Cake Precipitation

While a conventional process would have at this stage a solvent extraction or ion exchange process, such a high grade leach filtrate needs not, of course, an upgrading stage. However, some impurities must be removed as well as the excess of acid before precipitating a marketable yellowcake.

This step is done by raising the pH to 3.2 with milk of lime causing calcium sulfate and ferric hydroxide to precipitate out of the solution. The solids are then allowed to settle in 24'Ø iron cake thickener and filtered in a 2.5m^2 horizontal belt filter. The filtrates are recycled to the iron cake thickener, the overflow of which is sent to the yellowcake precipitation.

The iron cake discharge from the filter is then leached in a mild sulfuric acid at pH 2.8 to recover any uranium that might have been precipitated with the calcium sulfate and ferric hydroxide. Typical results show that between 1 and 2%

uranium are found in the iron cake at a precipitation pH of 3.2 and 75% of this uranium is recovered by the iron cake leach circuit. The final iron cake is then filtered and washed on 2.5m^2 horizontal belt filters and discharged to the effluent treatment section.

1.7 Yellowcake Precipitation and Drying (ref. figure 6)

From the purified high grade uranium liquor, the uranium is precipitated at pH 7 with a 10% solid milk of magnesia. This is done at room temperature in 3 agitated vessels providing a little over 3 hours retention time. The yellowcake slurry is then thickened in a 20'Ø yellowcake thickener yielding a 15% solids yellowcake slurry in the underflow. Part of the thickener underflow is recycled to the yellowcake precipitation tanks to provide seeds for the magnesium diuranate and ensure a total reaction of the magnesium hydroxide. This allows the final yellowcake to contain less than 2.5% magnesium.

The yellowcake thickener underflow is sent to a 7.5m^2 horizontal belt filter for filtration and production of a filter cake at about 45% moisture. The cake is gravity discharged to an oil fired tunnel type Proctor and Schwartz dryer operating at 375°F. The final product contains less than 2% moisture while the yellow cake dust is recovered in a venturi type wet dust collector (Kinpactor from American Air Filter).

The final product is stored in either of two 15 ton bins and packaged in a conventional manner in 55 gallon drums.

1.8 Effluent Treatment

The above described process yields 3 different types of residues:

a) the gravimetric tails, grading around 0.8% uranium which are stored separately for processing at a later date along with lower grade ores from other orebodies.

b) the leach residues representing about 50% of the weight of the solid leach feed and containing more than 99% of the radium 226. Overall this will represent some 5000 Tonnes of residues containing some 300,000 picocuries of radium per gram (with a conventional ore the leach residues contain around 1000 picocuries of radium per gram of solid). For environmental and radioprotection reasons residue is stored in concrete vaults.

c) The iron cake and other liquid effluent containing about 1500 picocuries of radium 226 per liter is treated with Barium chloride then with lime to raise the pH to 8.5. Part of the liquid is recycled in the process while a 5% slurry is sent to a tailings pond. The liquid discharge of this pond is further treated with Barium chloride, filtered in sand filters and the barium radium sulfate coprecipitate is allowed to settle in 2 trains of settling pond operated in parallel.

2. PROBLEMS ENCOUNTERED IN PROCESSING HIGH GRADE ORES

Processing very high grade ore is very appealing but unique problems were encountered by the designers and the operators and must be dealt with in a special way. Some problems are of a process nature while others are due to the radiation resulting from the high radium content in some parts of the mill.

2.1 Problems related to the Process

Like many other uranium ores in the Athabasca basin, the gangue is relatively acid soluble especially for the very fine fraction. This means that leaching must be very carefully controlled in order to avoid the dissolution of too many impurities. The controls include temperature, oxidation potential and conductivity. Operators had to be trained to overcome the traditional "the more, the better" syndrome.

It must be realized that the iron cake precipitation step, although efficient to remove iron and excess acidity, does not remove at a large extent impurities like molybdenum and vanadium like a solvent extraction or ion exchange would do. It follows that the produced yellow cake contains more of these impurities than the refiners like to see. Lately another impurity has been unexpectedly troublesome, the fluorine, which is present in the deeper zones of the D orebody.

Another process related problem is due to the high specific gravity of the ore and slurries that the mill has to deal with. A 30% uranium pitchblend slurry has a high solid specific gravity so that agitation in surge tanks and thickener must be carefully designed to avoid sanding out.

A third problem is related to the size of the equipment. Grinding and leaching for example have a capacity of 1.2 T/h. This is the size of a large pilot plant and the pump and pipe market had to be searched thoroughly to find reliable pumps for example, able to handle such a slurry. After several trials we have selected air operated diaphragm pumps.

2.2 Problems related to the radiation hazards

To put this problem into perspective it must be reminded that at contact a 30% grade ore has a gamma radiation around 80 to 100 mrad/hr compared to a 1 mrad/hr for a 0.3% grade and an overall allowable dose for the workers of 2.5 mrem/hr. It is then obvious that shielding has to be provided.

This is the reason why the storage bins, the concentrate thickener and the leach tanks are made of 12 to 18 inches thick concrete.

This alone allows to largely reduce the gamma radiation to an acceptable level. However, every time there is a spillage in this area the gamma level tends to increase. For that reason temporary restricted zones and temporary shielding have to be installed until the spillage is cleaned.

Another problem associated with the high grade ore is the radon concentration inside the mill. Again this affects especially the leach and filtration areas since the agitators and temperature helps the radon to emanate outside the leach tanks. Of course it must be realized that the quantity of radon emanated from this ore is not different from what is observed in a conventional mill because the total quantity of radon produced depends upon the quantity of radium present in the mill which in turn depends upon the quantity of uranium inventory in the mill (assuming equilibrium between uranium and radium).

However, since this radon emanates in a small area it may create some relatively high local concentration. Special ventilation techniques had then to be designed and, especially in winter, air changes had to be increased to 6 per hour in order to maintain a low radon concentration in the leach zone. This has been rather successful since in 15 months of operation, the mill had to be evacuated only twice because of excessive radon working levels.

TABLE 1

D ORE TYPICAL CHEMICAL ANALYSIS

	"High Grade"	"Low Grade"
U	22%	4%
Fe	4	6
Al	5	9
Si as SiO_2	45	58
Pb	2.7	0.45
Stot	0.3	0.3
Ctot	1.2	0.2
Mo	0.13	0.10
V	0.16	0.12
As	<0.01	<0.01

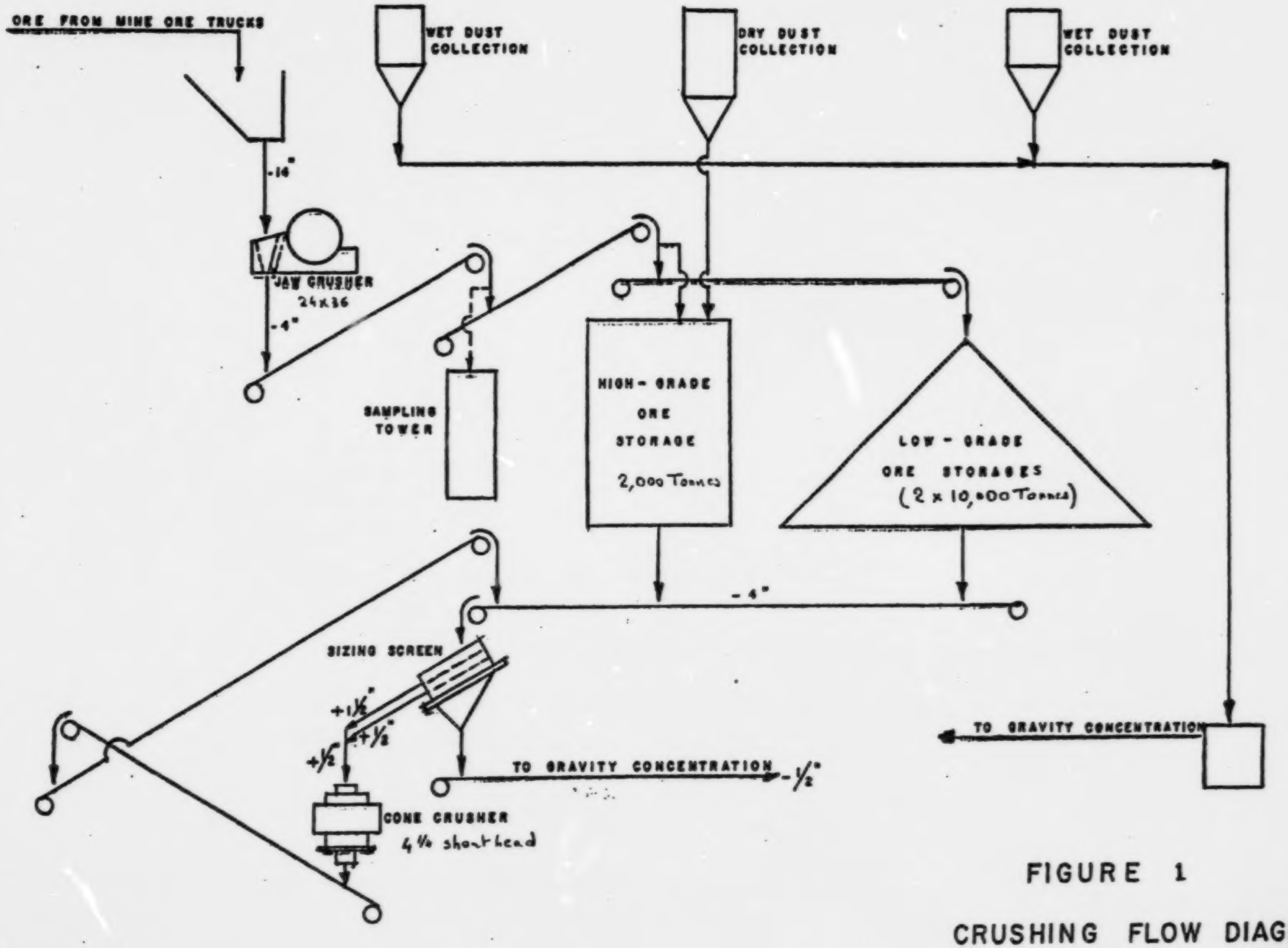

ORE FROM MINE ORE TRUCKS
WET DUST COLLECTION
DRY DUST COLLECTION
WET DUST COLLECTION
-14"
JAW CRUSHER
24x36
-4"
SAMPLING TOWER
HIGH - GRADE ORE STORAGE
2,000 Tonnes
LOW - GRADE ORE STORAGES
(2 x 10,000 Tonnes)
-4"
SIZING SCREEN
+1½"
+½"
-1½"
+½"
TO GRAVITY CONCENTRATION
-½"
CONE CRUSHER
4¼ short head
TO GRAVITY CONCENTRATION

FIGURE 1

CRUSHING FLOW DIAGRAM

TABLE 2

GRAVIMETRIC SEPARATION

Typical Screen Analysis of the Feed Ore

SIZE	% RETAINED (Cumulative)				
3/8"	3%	)			
3 mesh	24	)	Feed to the		
4 mesh	32	)	Yuba Jig		
6	43	)			
8	54	]			
10	63	]			
14	68	]			
20	73	]	Feed to the	)	
28	78	]	Denver Jig	)	Feed to the
35	82	]		)	Sand Tables
48	85	]		)	
65	89	]		]	
100	92	]		]	Feed to the
200	96	]		]	Fine Tables
pan	100		Cycloned		

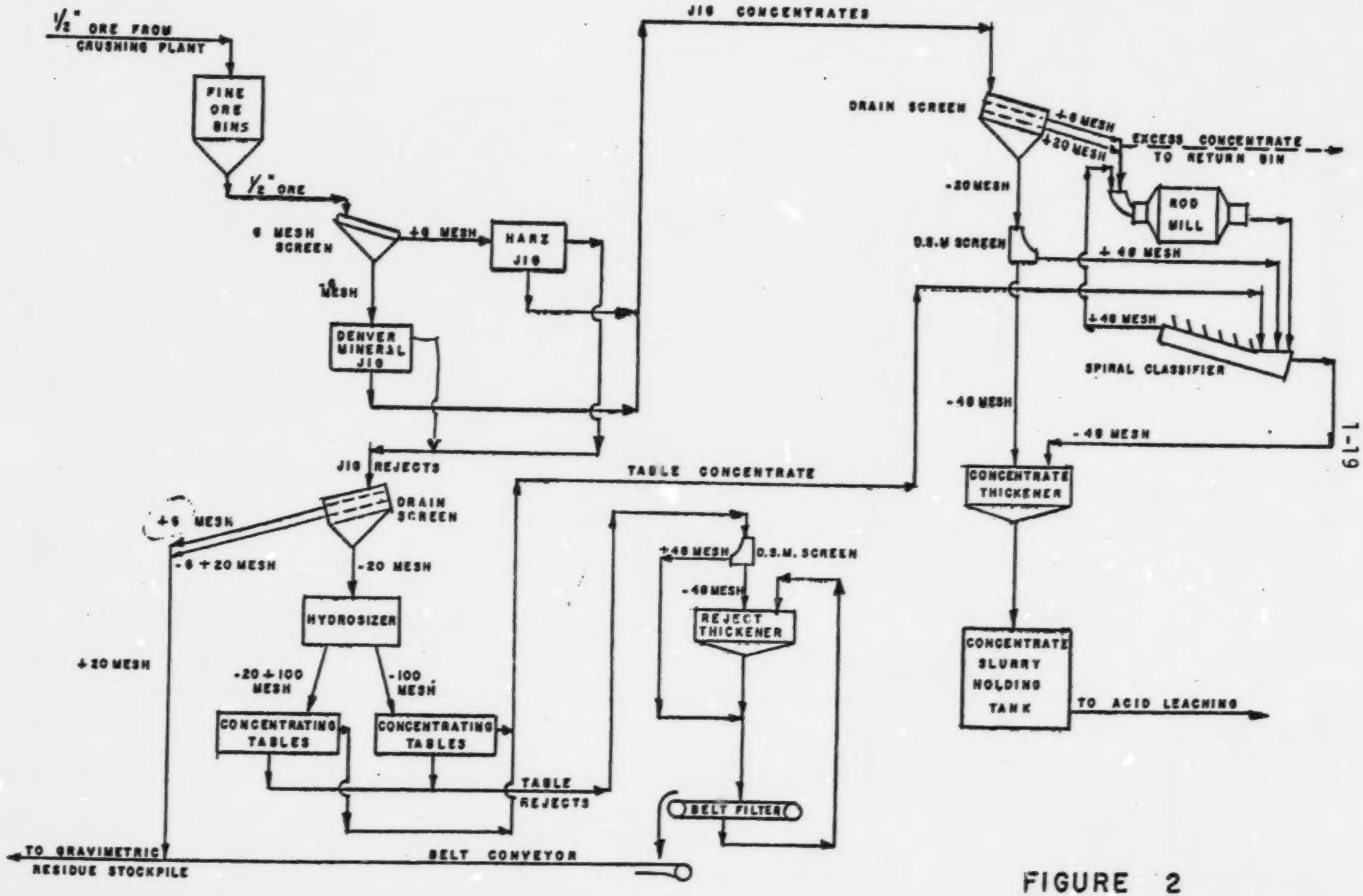
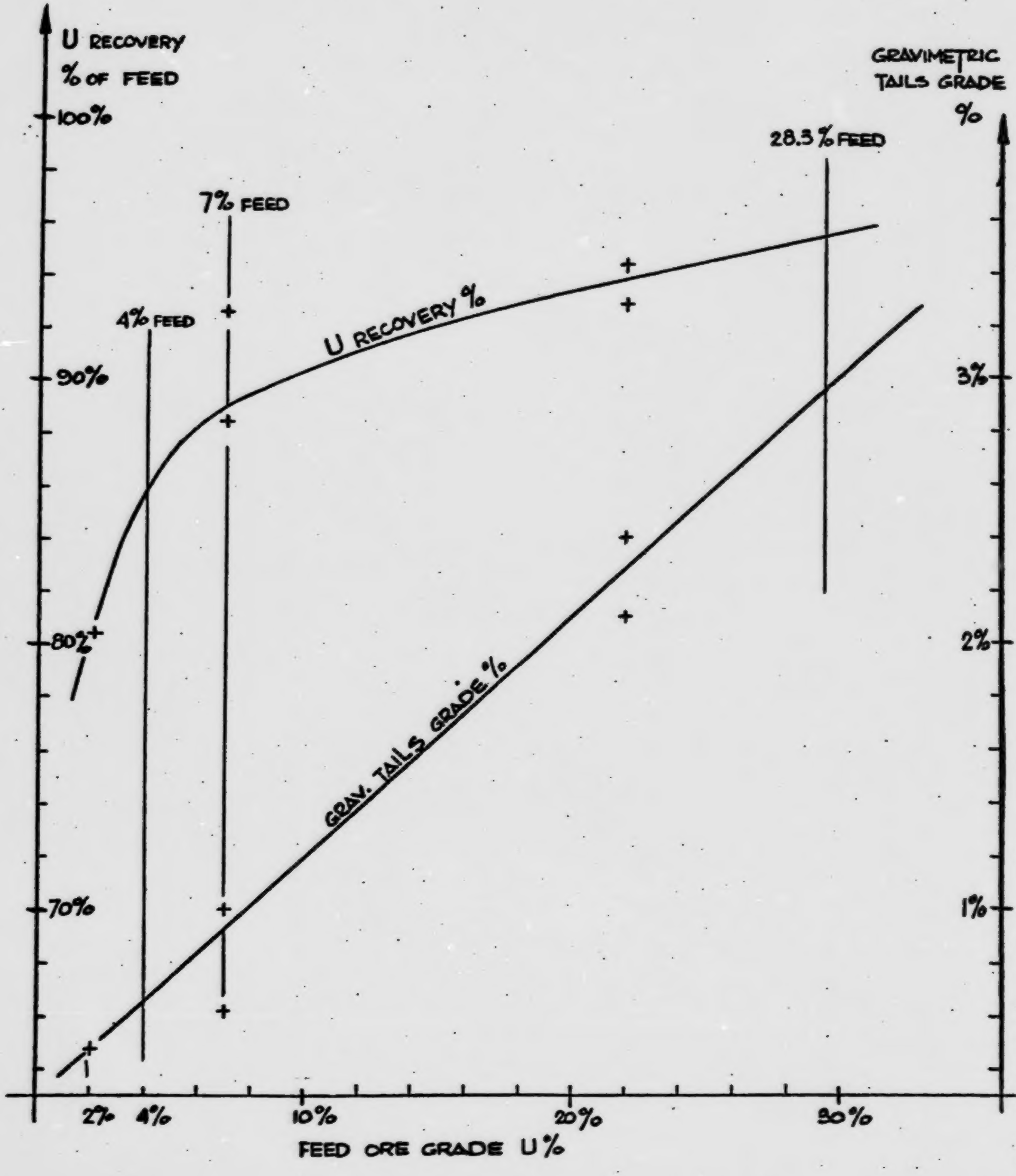

1/2" ORE FROM CRUSHING PLANT
FINE ORE BINS
1/2" ORE
6 MESH SCREEN
+6 MESH
-6 MESH
HARZ JIG
DENVER MINERAL JIG
JIG CONCENTRATES
JIG REJECTS
DRAIN SCREEN
+6 MESH
-6 + 20 MESH
+20 MESH
-20 MESH
HYDROSIZER
-20 + 100 MESH
-100 MESH
CONCENTRATING TABLES
CONCENTRATING TABLES
TABLE CONCENTRATE
TABLE REJECTS
+48 MESH
D.S.M. SCREEN
-48 MESH
REJECT THICKENER
BELT FILTER
BELT CONVEYOR
TO GRAVIMETRIC RESIDUE STOCKPILE
DRAIN SCREEN
+6 MESH
+20 MESH
EXCESS CONCENTRATE TO RETURN BIN
-20 MESH
ROD MILL
D.S.M SCREEN
+48 MESH
+48 MESH
SPIRAL CLASSIFIER
-48 MESH
-48 MESH
CONCENTRATE THICKENER
CONCENTRATE SLURRY HOLDING TANK
TO ACID LEACHING

FIGURE 2
GRAVITY CONCENTRATION
FLOW DIAGRAM

GRAVITY CONCENTRATION - PILOT PLANT RESULTS.

FIG 3



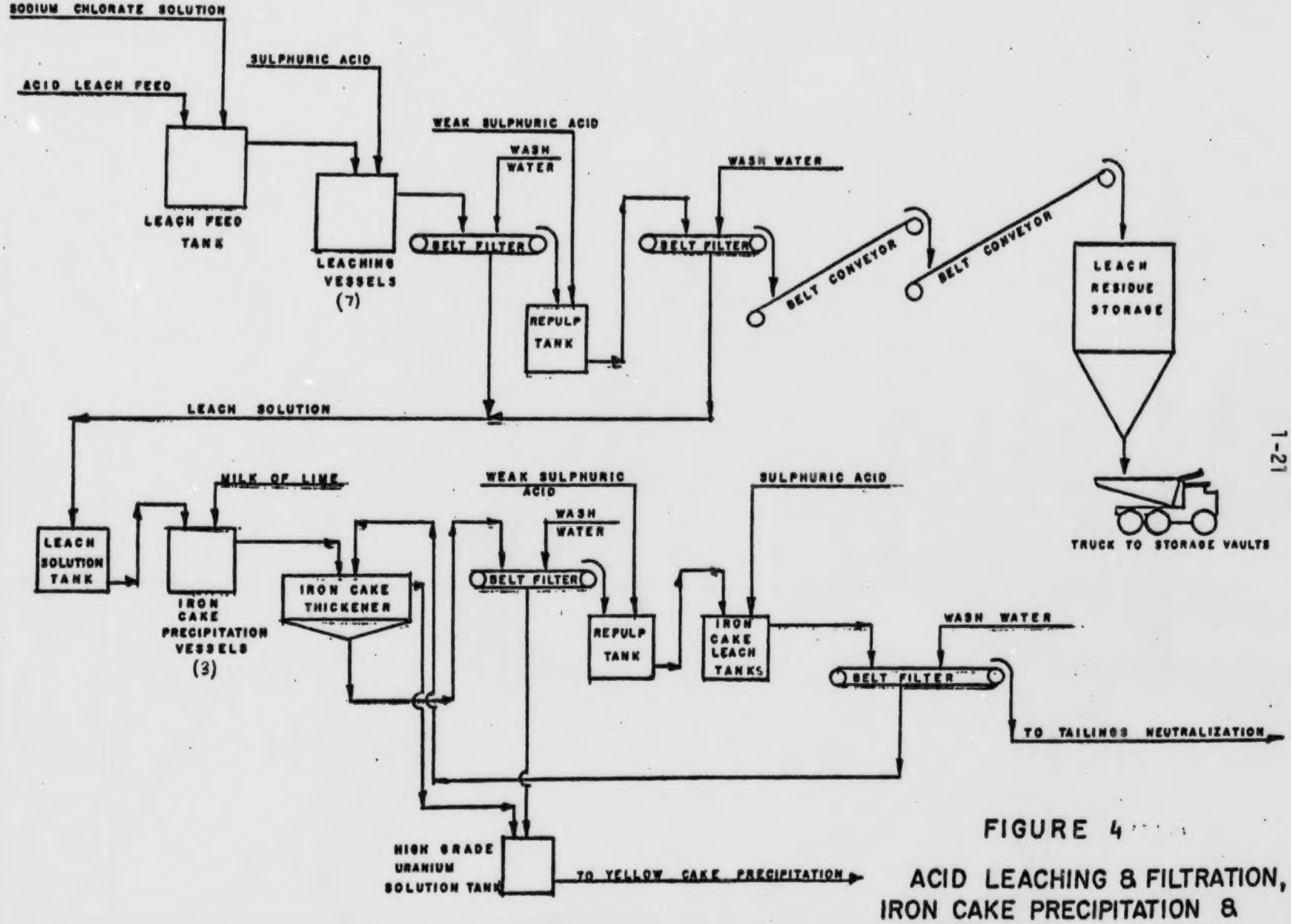

SODIUM CHLORATE SOLUTION
SULPHURIC ACID
ACID LEACH FEED
LEACH FEED TANK
LEACHING VESSELS (7)
WEAK SULPHURIC ACID
WASH WATER
BELT FILTER
REPULP TANK
WASH WATER
BELT FILTER
BELT CONVEYOR
BELT CONVEYOR
LEACH RESIDUE STORAGE
TRUCK TO STORAGE VAULTS
LEACH SOLUTION
MILK OF LIME
WEAK SULPHURIC ACID
WASH WATER
SULPHURIC ACID
LEACH SOLUTION TANK
IRON CAKE PRECIPITATION VESSELS (3)
IRON CAKE THICKENER
BELT FILTER
REPULP TANK
IRON CAKE LEACH TANKS
WASH WATER
BELT FILTER
TO TAILINGS NEUTRALIZATION
HIGH GRADE URANIUM SOLUTION TANK
TO YELLOW CAKE PRECIPITATION

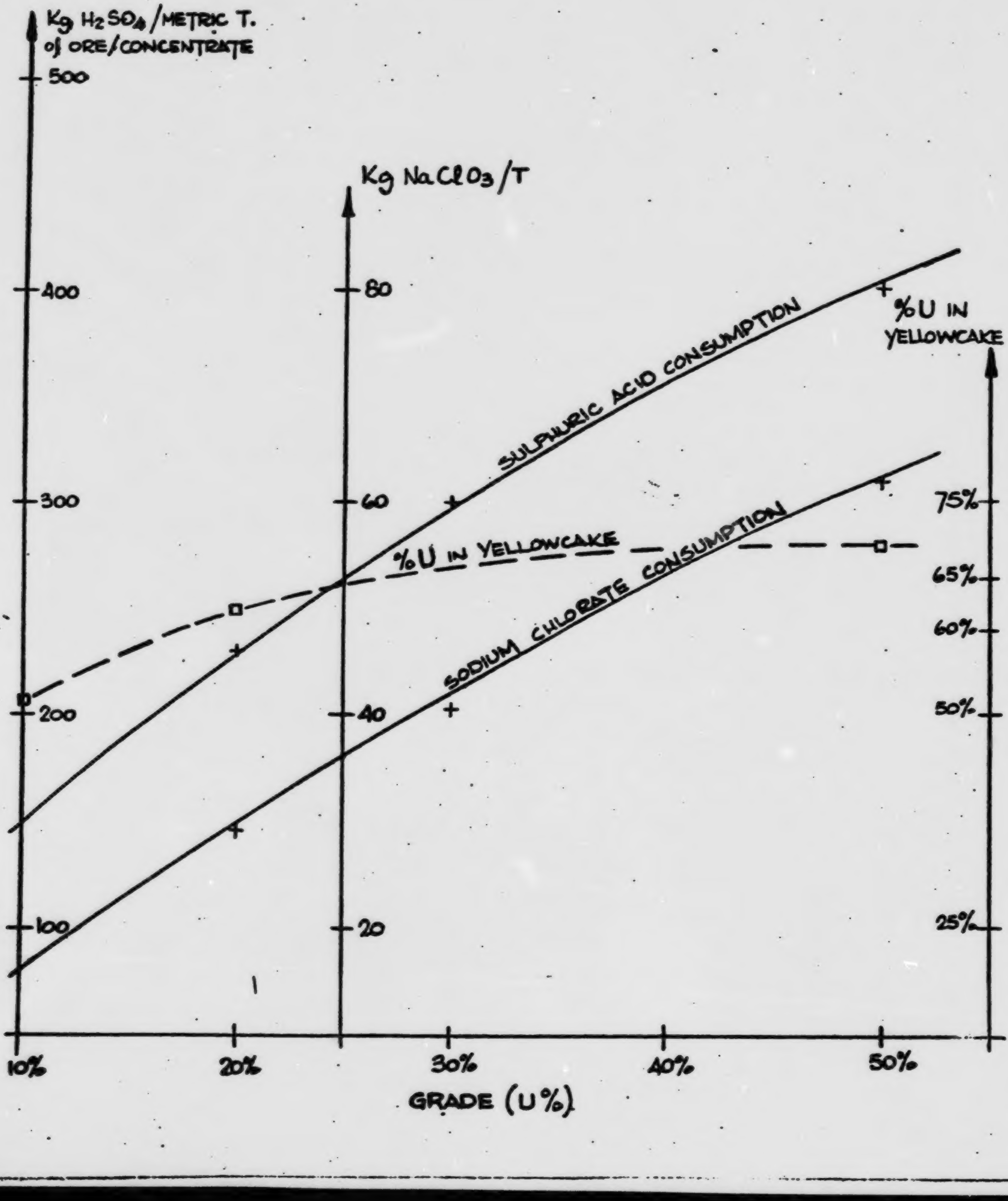
FIGURE 4

ACID LEACHING & FILTRATION, IRON CAKE PRECIPITATION & FILTRATION FLOW DIAGRAM

FIG 5

ACID LEACH

REAGENT CONSUMPTION VS. ORE GRADE



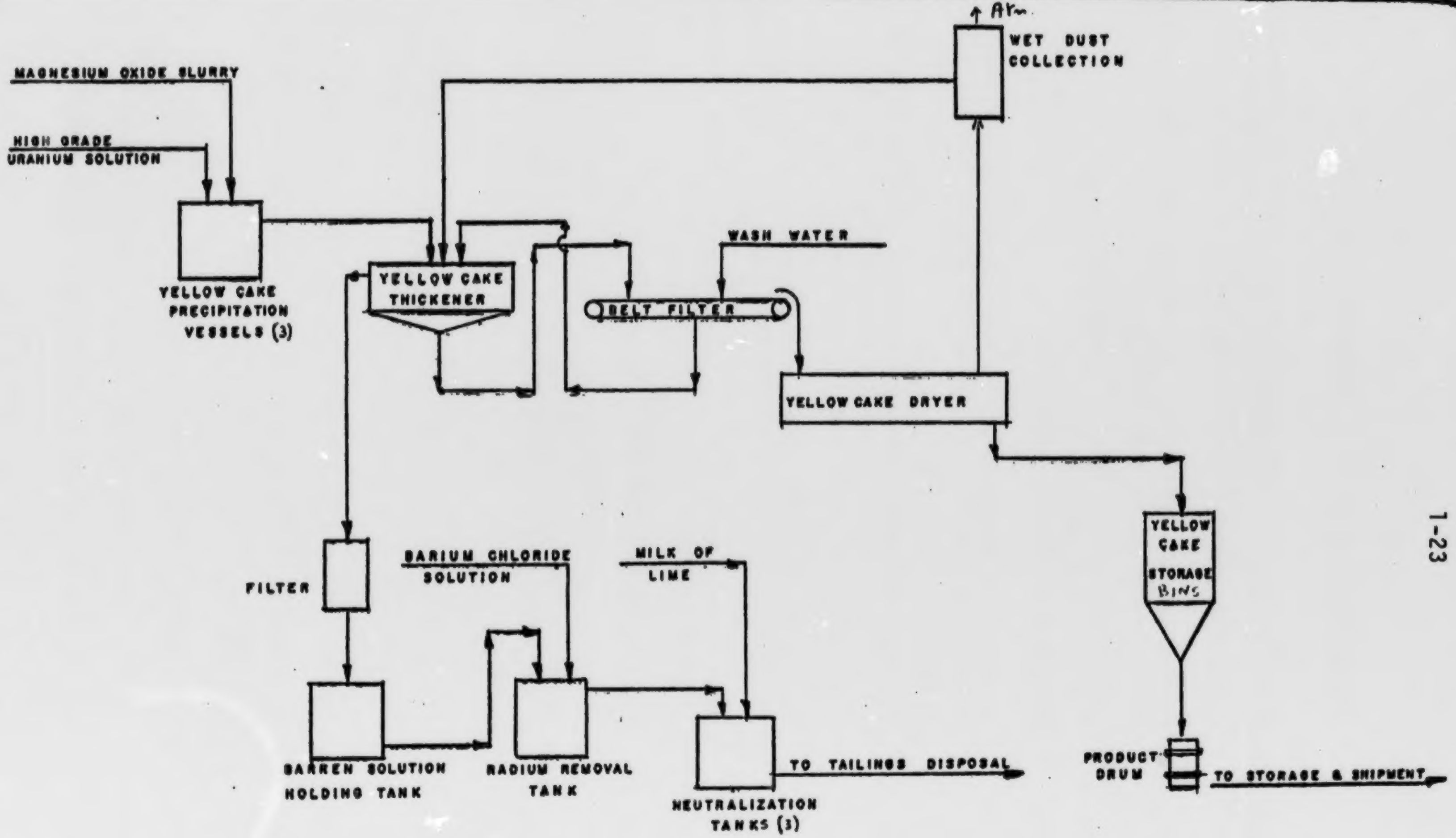

Air
WET DUST COLLECTION
MAGNESIUM OXIDE SLURRY
HIGH GRADE URANIUM SOLUTION
YELLOW CAKE PRECIPITATION VESSELS (3)
YELLOW CAKE THICKENER
WASH WATER
BELT FILTER
YELLOW CAKE DRYER
YELLOW CAKE STORAGE BINS
FILTER
BARIUM CHLORIDE SOLUTION
MILK OF LIME
BARREN SOLUTION HOLDING TANK
RADIUM REMOVAL TANK
NEUTRALIZATION TANKS (3)
TO TAILINGS DISPOSAL
PRODUCT DRUM
TO STORAGE & SHIPMENT

FIGURE 6

YELLOW CAKE PRECIPITATION, FILTERING & DRYING & TAILINGS NEUTRALIZATION & DISPOSAL FLOW DIAGRAM

DISCUSSION OF PAPER BY P. FOURNIER

L. Melis	With the high molybdenum levels, were there any problems in the tailings and what is the molybdenum solubility during the leach?
P. Fournier	We have exceeded the levels in the tailings. The dissolution of molybdenum is as high as 80%.
D. Rollwagen	What is the silica content in the yellowcake?
P. Fournier	No analysis are available.
V. Lakshmanan	How many air turnovers are there, and what is the radon content? Are there any amenability test procedures with respect to the product?
P. Fournier	The radon measures 4 millerehms in the leaching area of the mill. The product occasionally did not meet the amenability tests probably due to overflocculation.
M. Raghavayya	What is the depth of cover over the waste? What is the radioactive levels in the tailings?
P. Fournier	We cover the waste with one meter of soil. There are several types of rejects: a) gravity rejects (0.6-0.8% U_3O_8) b) leach tails - contains 99.9% Ra^{226}. There is 1000-2000 pCi/g Ra^{226} in the tails so they are stored in concrete containment areas. c) iron cake d) neutralized overflow, which is sent to the regular tailings system. It is essentially $CaSO_4$. The effluent is treated with $BaCl_2$.
J. Goode	Are you controlling the leach by on-line analysis?
P. Fournier	Conductivity is being used. No on-line analyses are used.
C. Lendrum	Are there any trucking problems?
P. Fournier	The tails are discharged by belt.
V. Lakshmanan	Do you contemplate any molybdenum separation eventually?
P. Fournier	No, we plan to control the molybdenum problem during leaching.

For presentation at the Annual Meeting of Canadian Uranium Producers' Metallurgical Committee, Bancroft, Ontario, Canada 19-20 May 1982

CARO'S ACID - A NEW OXIDANT FOR ACID LEACHING OF URANIUM ORES

R.J. Ring,[*] D.J. Waters[†] and M. Yamine[†]

[*]Australian Atomic Energy Commission
Research Establishment
Lucas Heights Research Laboratories
NSW, 2232, Australia

[†]Interox Chemicals Pty Ltd
Botany, NSW, 2019, Australia

ABSTRACT

An experimental program has been carried out by the Australian Atomic Energy Commission to compare the effectiveness of Caro's acid as an alternative oxidant to pyrolusite in the leaching of uranium ores from the Alligator Rivers region of northern Australia. Batch leaching tests on Ranger, Nabarlek, Koongarra and Jabiluka ores showed that uranium extraction was unaffected by choice of oxidant, but that Caro's acid reduced acid requirements by 15-25%. Lime requirements for neutralisation of tailings liquors were reduced by 20-30%.

The batch studies were extended to a four-stage continuous leaching circuit with a capacity of 400 g ore h^{-1}. Comparative leaching tests confirmed the reagent savings obtained in the batch tests.

Following these encouraging laboratory results, Interox Chemicals Pty Ltd carried out plant trials using Caro's acid. Reagent savings in the mill were similar to, or better than, those obtained in laboratory testwork. The trials also established that Caro's acid has a number of operational advantages over pyrolusite.

1. INTRODUCTION

Uranium is most commonly extracted from its ores by leaching in sulphuric acid. Simple hexavalent uranium oxides and compounds are readily soluble but uranium in the tetravalent form is insoluble and must be oxidised for dissolution to occur. The most commonly used oxidants have been sodium chlorate and manganese dioxide. They act indirectly by maintaining iron as the ferric ion which oxidises uranium by a surface electrochemical reaction. An Fe^{3+} concentration of about 1-2 g L^{-1} is usually considered adequate for the effective dissolution of uraninite [1].

The choice between sodium chlorate and pyrolusite has usually been made on the basis of local availability since this affects delivered price. In North America and France, sodium chlorate is generally used, while pyrolusite has been preferred in Australia and South Africa. Both oxidants have disadvantages. When pyrolusite is used, the raffinate contains a high concentration of manganese (0.8-1.5 g L^{-1}) which is a potential heavy metal pollutant. Even when the raffinate is treated with lime to pH 8, 100-200 mg Mn L^{-1} remains in solution [2]. However, manganese levels in the tailings dam decrease on prolonged storage to about 10-50 mg L^{-1}.

When sodium chlorate is used, there is a chloride build-up in the mill circuits. Water management policy for the new Australian mills in the Alligator Rivers region of the Northern Territory requires zero tailings water release which means that water in the tailings dam must be recycled through the mill. Under this system, chloride concentration could increase to the level where it would decrease the efficiency of solvent extraction. High chloride levels are also objectionable on the grounds that they mobilise radium [3].

Permonosulphuric acid (Caro's acid - H_2SO_5), a derivative of hydrogen peroxide, has a number of advantages over pyrolusite as an oxidant in leaching applications. It is attractive on environmental grounds because the residual reduction product is water. Because it is a liquid, storage, handling and metering are simplified, and a grinding circuit is not required. Although the unit cost of Caro's acid reagent is higher than that of pyrolusite, its use will result in reductions in acid consumption, and subsequently lime for neutralisation.

A limited study of the behaviour of peroxygen oxidants in the acid leaching of uranium ores was conducted by Burkin and Monhemius [4]. For Elliot Lake and Rabbit Lake ores, Haque and Ritcey [5] showed that peroxygen compounds were as effective as commonly used oxidants.

This paper presents results of laboratory research at the AAEC Research Establishment to compare the effectiveness of Caro's acid and pyrolusite in the leaching of ores from the Alligator Rivers region. Following encouraging results, Interox Chemicals constructed a plant to generate Caro's acid and carried out trials at the Queensland Mines uranium mill at Nabarlek in the Northern Territory. The findings of the plant trials are also discussed.

2. PROPERTIES OF PEROXYGEN OXIDANTS

Hydrogen peroxide, the cheapest source of active oxygen, is generally an effective oxidant in both acid and alkaline media. However, hydrogen peroxide tends to decompose catalytically in the presence of metal ions such as copper and iron :

$$2H_2O_2 \longrightarrow 2H_2O + O_2 \qquad (1)$$

However, Caro's acid has a greater stability to catalytic decomposition and is preferred in leaching applications. It can be made in a number of ways but, for economic reasons, the preferred method is by controlled addition of hydrogen peroxide to concentrated sulphuric acid with cooling to remove heats of dilution and reaction :

$$H_2SO_4 + H_2O_2 \rightleftharpoons H_2SO_5 + H_2O \quad . \tag{2}$$

In a uranium processing application, only H_2O_2 needs to be transported to the mill as sulphuric acid is already required for leaching. A high yield of Caro's acid is favoured by excess sulphuric acid and high concentrations of reagents. For uranium leaching, Caro's acid is normally made as 25 wt % H_2SO_5, using 98 wt % H_2SO_4 and 50-70 wt % H_2O_2 at a mole ratio, $H_2SO_4 : H_2O_2$, of about 3:1. A typical product composition (wt %) is 26% H_2SO_5, 1% H_2O_2, 57% H_2SO_4 and 16% H_2O, which should be kept at 10-20°C to avoid excessive decomposition (half life 10 days at 40°C) [6].

$$2H_2SO_5 \longrightarrow 2H_2SO_4 + O_2 \tag{3}$$

In practice, Caro's acid is made by a continuous plant located on site, with a flexible rate of production tied to rate of consumption, so that storage is unnecessary.

Equations 4-7 show that Caro's acid is a stronger oxidising agent than hydrogen peroxide and other oxidants commonly used in uranium ore leaching. (E_o potentials are relative to standard hydrogen electrode).

$$HSO_5^- + 2H^+ + 2e^- \longrightarrow HSO_4^- + H_2O \quad E_o = 1810 \text{ mV [7]} \tag{4}$$

$$H_2O_2 + 2H^+ + 2e^- \longrightarrow 2H_2O \quad E_o = 1780 \text{ mV} \tag{5}$$

$${}^1/_3\, ClO_3^- + 2H^+ + 2e^- \longrightarrow {}^1/_3\, Cl^- + H_2O \quad E_o = 1450 \text{ mV} \tag{6}$$

$$MnO_2 + 4H^+ + 2e^- \longrightarrow Mn^{2+} + 2H_2O \quad E_o = 1210 \text{ mV} \tag{7}$$

$$2Fe^{3+} + 2e^- \longrightarrow 2Fe^{2+} \text{ (sulphate)} \quad E_o = 640 \text{ mV [8]} \tag{8}$$

Equations 4 and 7 predict that acid consumed in oxidation by manganese dioxide is twice that for Caro's acid.

In an oxidation reaction, the reduction of Caro's acid (equation 4) produces sulphuric acid equivalent to that required for its formation. For convenience, in this paper, consumption of Caro's acid is expressed as H_2O_2 equivalent; on the basis of equivalent weights, reagent requirements per g Fe^{2+} oxidised are as in Table 1.

TABLE 1

COMPARISON OF REAGENT CONSUMPTIONS FOR OXIDATION

(per g Fe^{2+})

Oxidant	Oxidant Consumption (g)	Acid Consumption (g)
Manganese dioxide	0.777	1.75
Pyrolusite (79% MnO_2)	0.983	1.75*
Sodium chlorate	0.317	0.875
Caro's acid	0.304 (as H_2O_2)	0.875
Hydrogen peroxide	0.304	0.875

*Assumes other minerals in pyrolusite do not dissolve under acid conditions

3. COMPARISON OF CARO'S ACID AND PYROLUSITE

Since pyrolusite was originally chosen as the oxidant for the leaching of ores from the Alligator Rivers region, it was adopted as a standard for the assessment of Caro's acid. The experimental program consisted of pairs of experiments conducted under identical conditions to compare reagent consumptions and uranium extractions. The ratio of oxidant consumptions and reagent savings were used as a measure of performance. These are defined below :

R_{OX} (oxidant ratio) $= P/C$

S_{ACID} (acid saving) $= A_p - A_c$

$\% S_{ACID}$ (per cent acid saving) $= (S_{ACID}/A_p)100 = (1 - A_c/A_p)100$

S_{LIME} (lime saving) $= L_p - L_c$

$\% S_{LIME}$ (per cent lime saving) $= (S_{LIME}/L_p)100 = (1 - L_c/L_p)100$

where P = pyrolusite consumption (kg t^{-1} ore)

C = Caro's acid consumption, as H_2O_2 (kg t^{-1} ore)

A = acid consumption (kg t^{-1} ore)

L = lime ($Ca(OH)_2$) consumption (kg t^{-1} ore)

Subscripts p and c refer to pyrolusite and Caro's acid respectively

Table 1 shows that the theoretical oxidant ratio (R_{OX}) is 3.2 for pyrolusite containing 79% MnO_2.

Equations 4 and 7 show that the acid saving (S_{ACID}) gained by the use of Caro's acid depends on the oxidant consumption. In theory from Table 2

$$S_{ACID}/P = (1.75 - 0.875)/0.983 = 0.89 \quad (9)$$

$$S_{ACID}/C = 0.89 \times 3.2 = 2.85 \quad (10)$$

The lime saving (S_{LIME}) is also a function of oxidant consumption. The residual reduction product of Caro's acid is water, whereas pyrolusite releases manganous ions which are almost completely (92-95%) precipitated at the neutralisation pH of 8.0. The lime required for removal of manganese can be determined from equation 11.

$$MnSO_4 + Ca(OH)_2 \longrightarrow Mn(OH)_2 + CaSO_4 \quad . \quad (11)$$

From the stoichiometry of equation 11

$$S_{LIME}/P = 0.673 \quad . \quad (12)$$

3.1 Experimental Plan

The comparative laboratory experiments were carried out in three steps of increasing scale :

(1) The effectiveness of pyrolusite, hydrogen peroxide and Caro's acid were determined for a wide range of batch leaching conditions using Ranger ore. Duration of leaching was 8 h.

(2) Batch leaching tests of 24-36 h duration were carried out on all ores to compare reagent consumptions under typical or expected processing conditions. Lime requirements for neutralisation of leach liquor/tailings slurries were also determined.

(3) The batch testwork was extended to a continuous leaching circuit. The performance of Caro's acid and pyrolusite were compared under the operating conditions of the Ranger and Nabarlek mills.

4. ORE SAMPLES

Ore samples were obtained from the Ranger, Nabarlek, Koongarra and Jabiluka deposits in the Alligator Rivers region. These are all unconformity vein-type deposits in which the major host rocks are quartz, chlorite, feldspar and muscovite schists. The primary mineral is uraninite, with sklodowskite, saleeite and gummite the most common secondary minerals [9]. More detailed mineralogical data and uranium grades for ores used in this work are presented in Table 2.

TABLE 2

MINERALOGY OF URANIUM ORES[9]

ORE	GRADE(1) (% U_3O_8)	HOST ROCK AND GANGUE MINERALS	URANIUM MINERALS
Ranger	0.25-0.37	Primary Zone(2) Mainly quartz-chlorite schist with chert, granitic and basic dykes, chlorite rocks and dolomites. Oxidised Zone Kaolinite	Primary Zone Pitchblende, brannerite and uranium-titanium, -phosphate and -silicate minerals. Oxidised Zone Saleeite, sklodowskite, gummite and metatorbernite.
Nabarlek	1.6-1.9	Primary Zone(2) Mainly muscovite-chlorite schists with hematite Oxidised Zone Muscovite, chlorite, quartz and hematite/goethite with kaolinite	Primary Zone Pitchblende and uranium-titanium and uranium-silicate minerals. Oxidised Zone Sklodowskite, coffinite, curite, kasolite rutherfordine, uraniferous zircon and uranium-titanium minerals.
Koongarra	0.45-0.70	Primary Zone(2) Mainly quartz-chlorite-muscovite schist. Graphite and garnet are accessory minerals	Primary Zone Uraninite Oxidised Zone Sklodowskite, kasolite, renardite, metatorbernite, saleeite, curite and a lead-uranium phosphate mineral.
Jabiluka	0.53	Primary Zone(2) Quartz-chlorite-graphite-pyrite schists and quartz-chlorite-graphite-sericite-breccia with lesser amounts of calcite, kaolinite, montmorillonite, apatite, tourmaline, sphene, leucoxene and rutile. Native gold is also present.	Primary Zone Pitchblende and uraninite. Oxidised Zone Autunite, sklodowskite and saleeite.

(1) Samples used in this study.

(2) Pyrite, chalcopyrite, galena and various iron, manganese and titanium oxides/hydroxides are present in minor to trace amounts.

All the Alligator River ores require similar leaching conditions. Although leaching is relatively straightforward, it is important to select the best combination of pH, temperature and leaching time to optimise the consumption of acid against recovery of uranium [10]. The reactivity of mica and chlorites, in particular, is very sensitive to pH within the range required for dissolution of uraninite. In general, ore ground to 100% less than 300 µm and 50% less than 75 µm is leached as a 45-55 wt % slurry for 20-36 h at pH 1.6-2.0, 35-45°C and an oxidation potential of 425-475 mV.* Sulphuric acid is added to all or most of the leaching vessels to control either pH or free acid while oxidant is generally added for the first 12 h of leaching.

Pyrolusite (79% MnO_2) from Groote Eylandt was ground to 100% less than 150 µm and 80% less than 75 µm for use in leaching experiments.

5. BATCH TESTWORK

5.1 Experimental Equipment and Procedures

Leaching tests were carried out in baffled, stainless steel, 600 mL beakers. In each test, 390 g of ore was used at an initial slurry density of 55 wt %. During leaching, the volume of solution was increased by the addition of sulphuric acid, typically 10-13 mL. Slurries were stirred by four-bladed impellers operating at 5 rev s^{-1}. Temperature was controlled to within ±0.5°C by a water bath. Redox potential and pH were monitored continuously and controlled by additions of A.R. grade sulphuric acid and oxidant. In some experiments, oxidant addition was stopped after 12 h leaching.

Caro's acid was sampled immediately before each experiment and analysed by the recommended method [6]. The Caro's acid compositions used were as-made (25 wt %) and diluted (7.7 and 10 wt %). During experiments, reagents were kept at 6-8°C to minimise decomposition.

The progress of leaching was followed by withdrawing 15 mL slurry samples at appropriate intervals. The samples were filtered and the residues washed repeatedly and finally dried at 105°C. Leaching efficiencies were determined by assaying the dried solids for uranium by means of a delayed neutron analysis technique which has an accuracy of ±3%. A small correction was applied to account for ore that had dissolved during leaching. Filtrates were analysed for numerous metal ions.

Neutralisation tests were carried out on leach liquor/tailings slurries at 25°C after dilution to a slurry density of 40 wt % to simulate plant conditions. Uranium was not removed before neutralisation but a correction (0.43 kg $Ca(OH)_2$/kg U) was made to lime consumption data. Slurries were neutralised to pH 8 by incremental additions of A.R. grade calcium hydroxide. Additions were made at intervals of about 20 minutes to give a change of 0.2-0.5 pH. Total time for neutralisation was about 4 h. Slurry samples were withdrawn at pH 7 and 8. Filtrates were analysed for manganese and numerous metal ions.

5.2 Uranium Extraction

For all ores, uranium extraction was unaffected whether pyrolusite, hydrogen peroxide or Caro's acid was used as the oxidant. This result was expected as oxidants act indirectly by converting Fe^{2+} to Fe^{3+}. Typical results for Nabarlek ore are shown in Table 3.

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*All redox potentials quoted for leaching conditions are relative to a saturated calomel electrode at the leaching temperature. The most highly oxidised solution is assumed to have the most positive potential.

TABLE 3

URANIUM EXTRACTION FOR NABARLEK ORE (%)

(40°C, pH 1.7, 480 mV)

Duration of Leaching (h)	Oxidant		
	Pyrolusite	Caro's Acid	Hydrogen Peroxide
0.5	78.0	77.9	79.6
2	91.3	93.1	91.8
8	95.4	95.8	95.9
14	96.3	96.1	96.3
20	96.4	96.5	96.4
24	96.4	96.5	96.4

5.3 Comparison of Hydrogen Peroxide and Caro's Acid

When converted to Caro's acid, the consumption of hydrogen peroxide was 20-40% lower than for hydrogen peroxide itself. Results in Table 4 show that the consumption of hydrogen peroxide increased markedly at high temperatures or under strong oxidising conditions (potential >400 mV) probably because of decomposition reactions. The stability of Caro's acid was not affected by these conditions. Figure 1 compares H_2O_2 consumptions as a function of leaching time for Ranger ore under typical processing conditions.

TABLE 4

COMSUMPTIONS OF PEROXYGEN OXIDANTS

(Ranger Ore, pH 1.7, 8 h Ieaching)

Temperature (°C)	Potential (mV)	H_2O_2 Consumption (kg t^{-1} ore)	
		Caro's Acid	Hydrogen Peroxide
35	475	0.85	1.32
55	475	1.52	2.62
35	395	0.42	0.52
35	475	0.85	1.32
35	550	1.2	6.7

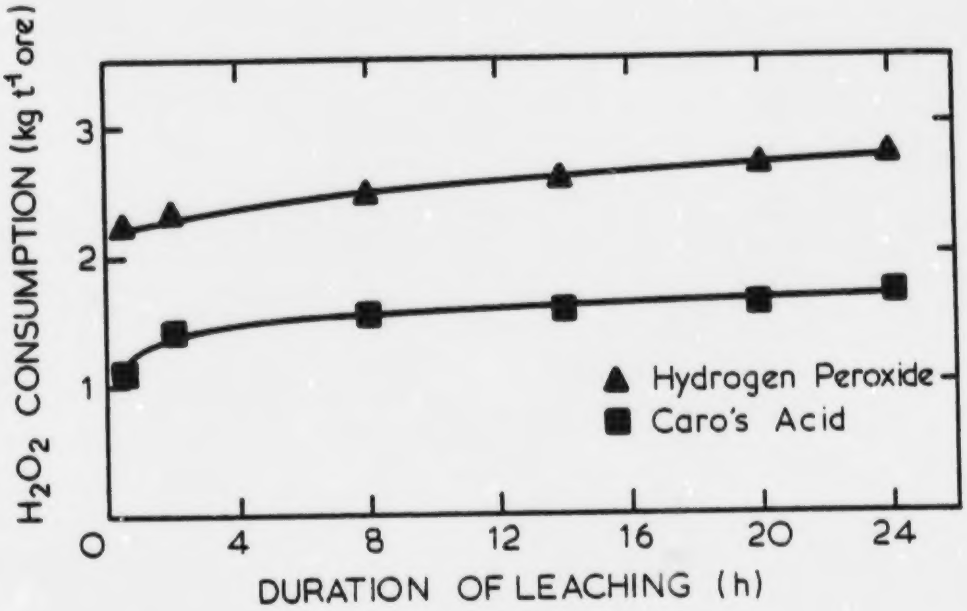

Figure 1. Oxidant consumptions for Ranger ore (35°C, pH 1.8, 475 mV)

5.4 Reagent Consumptions

The reagent savings shown in Table 5 were established for the use of Caro's acid as an alternative oxidant to pyrolusite.

TABLE 5

REAGENT SAVINGS FOR CARO'S ACID

Ore	Reduction in Acid Consumption (%S_{ACID})	Reduction in Lime Consumption (%S_{LIME})	Pyrolusite/H_2O_2 Consumption (R_{OX})
Ranger	15-28	18-25	3.3-3.9
Nabarlek	15-19	20	3.1-3.3
Koongarra	15-17	19	2.7-3.3
Jabiluka	16-20	20-29	2.5-2.9

Oxidant Ratio (R_{OX})

R_{OX} was typically very close to the stoichiometric value of 3.2 for

Ranger, Nabarlek and Koongarra ores, but less for Jabiluka. Standard procedure of controlling redox potential from the start of leaching was not suitable for Jabiluka ore as hydrogen sulphide was evolved by initial acid reactions with sulphides. Under these conditions, R_{OX} was 2.5. In a few tests with delayed oxidant addition, R_{OX} increased to 2.8-2.9. Leaching conditions were not optimised for Jabiluka ore and it is likely that R_{OX} could be further improved.

Acid Saving (S_{ACID})

Acid consumption curves for Nabarlek ore are shown in Figure 2. In theory, S_{ACID}/P should be 0.89. Separate experiments in which ferrous sulphate solution was oxidised by incremental additions of pyrolusite and Caro's acid showed that S_{ACID} was about 11% higher than this theoretical value. In leaching experiments, S_{ACID} was even greater. The average was given by

$$S_{ACID}/P = 1.25 \qquad (13a)$$

or

$$S_{ACID}/C = 4 \qquad (13b)$$

The acid that was unaccounted for could have been caused by a slight decrease in the dissolution of major gangue minerals when Caro's acid was used. Table 6 shows that leach liquors contained increased concentrations of iron, aluminium and magnesium for pyrolusite as oxidant. This can be attributed to the dissolution of impurities in the pyrolusite and increased reactivity of gangue minerals because of the higher concentration of sulphate or bisulphate ion in solution.

TABLE 6

CONCENTRATIONS OF MAJOR COMPONENTS IN LEACH LIQUORS

(pH 1.6-1.8, 35-40°C, 450-475 mV for 24 h)

Ore	Ranger		Koongarra	
Oxidant	Pyrolusite	Caro's Acid	Pyrolusite	Caro's Acid
Acid Consumption ($kg\ t^{-1}$ ore)	45.8	32.9	55.3	48.3
Oxidant Consumption ($kg\ t^{-1}$ ore)	6.6	1.68	5.37	1.63
Component	Concentration ($mg\ L^{-1}$)			
Al	3200	2590	2400	1470
Ca	230	190	610	650
Fe	3360	2920	3660	3230
Mg	2700	2300	3870	3750
Mn	3000	40	2420	170

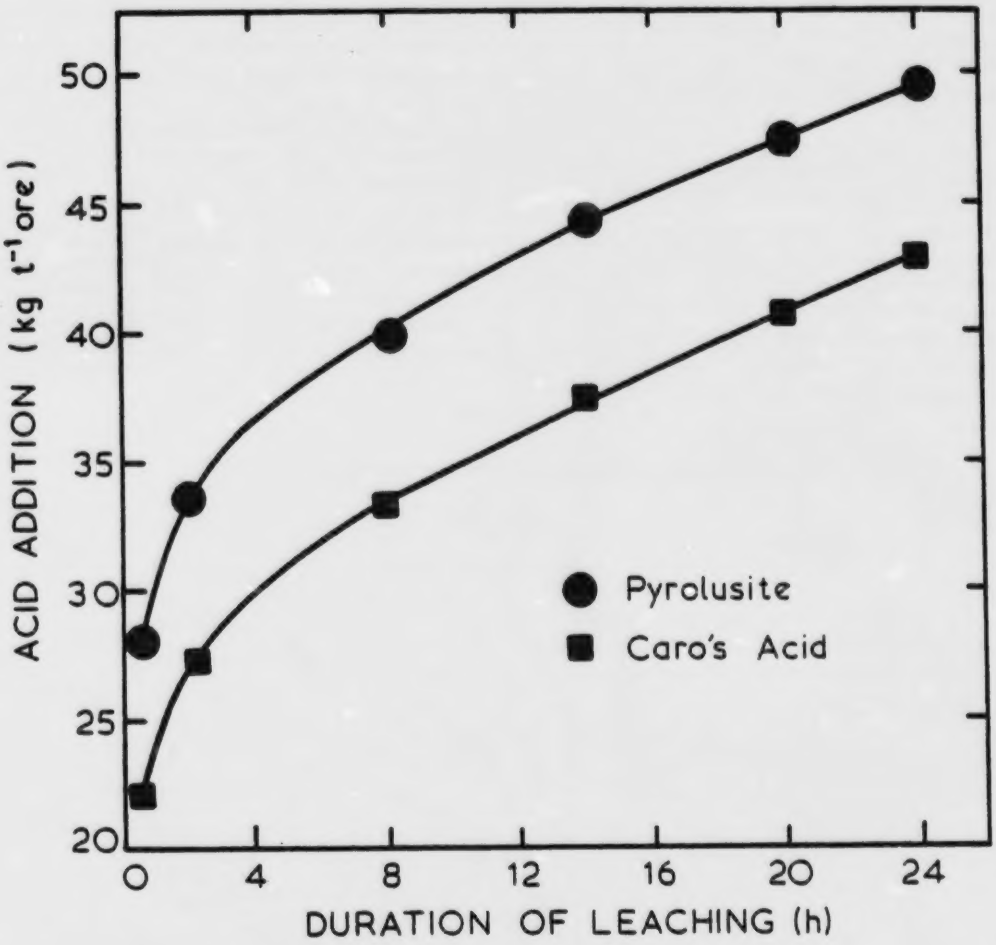

Figure 2. Acid consumption for Nabarlek ore (40°C, pH 1.7, 480 mV)

Lime Saving (S_{LIME})

Neutralisation curves for leach liquor tailings slurries derived from Koongarra ore are shown in Figure 3. For all ores, Table 5 shows that Caro's acid produced $\%S_{LIME}$ which were slightly greater than $\%S_{ACID}$. There is a theoretical reason why $\%S_{LIME} > \%S_{ACID}$. If a metal ion released in acid leaching is completely precipitated by lime, the ratio of lime/acid usage can be determined from the equation

$$H_2SO_4 + Ca(OH)_2 \longrightarrow CaSO_4 + 2H_2O \quad . \qquad (14)$$

This equation predicts a lime (as $Ca(OH)_2$)/acid usage of 0.76. This ratio should apply to most metal ions in the tails liquor, including manganese, since over 90% is precipitated at pH 8. The only important exception is magnesium which consumes little lime since it mostly remains in solution at pH 8. In practice then, the lime/acid requirements are less than 0.76. The neutralisation data indicate an average L_C/A_C ratio of 0.56 for Caro's acid as oxidant. When pyrolusite was used, the average L_P/A_P ratio was slightly higher (0.59) because the lime/acid ratio for manganese alone is close to 0.76.

The average S_{LIME} was about 40% greater than the value predicted from equation 12 because the use of Caro's acid also reduced the concentrations of ions other than Mn in the leach liquor. If these ions, and Mn, were completely precipitated in neutralisation, the stoichiometry of equation 14 predicts a maximum S_{LIME}/S_{ACID} of 0.76. The experimental average for S_{LIME}/S_{ACID} was 0.71. In practice, this ratio will vary depending on the proportion of extractable magnesium in the ore since magnesium is only partly precipitated at pH 8. In a mill application, S_{LIME}/S_{ACID} would be greater than 0.71 because of the lower reactivity of industrial lime and the less efficient reaction of lime in a continuous neutralisation circuit.

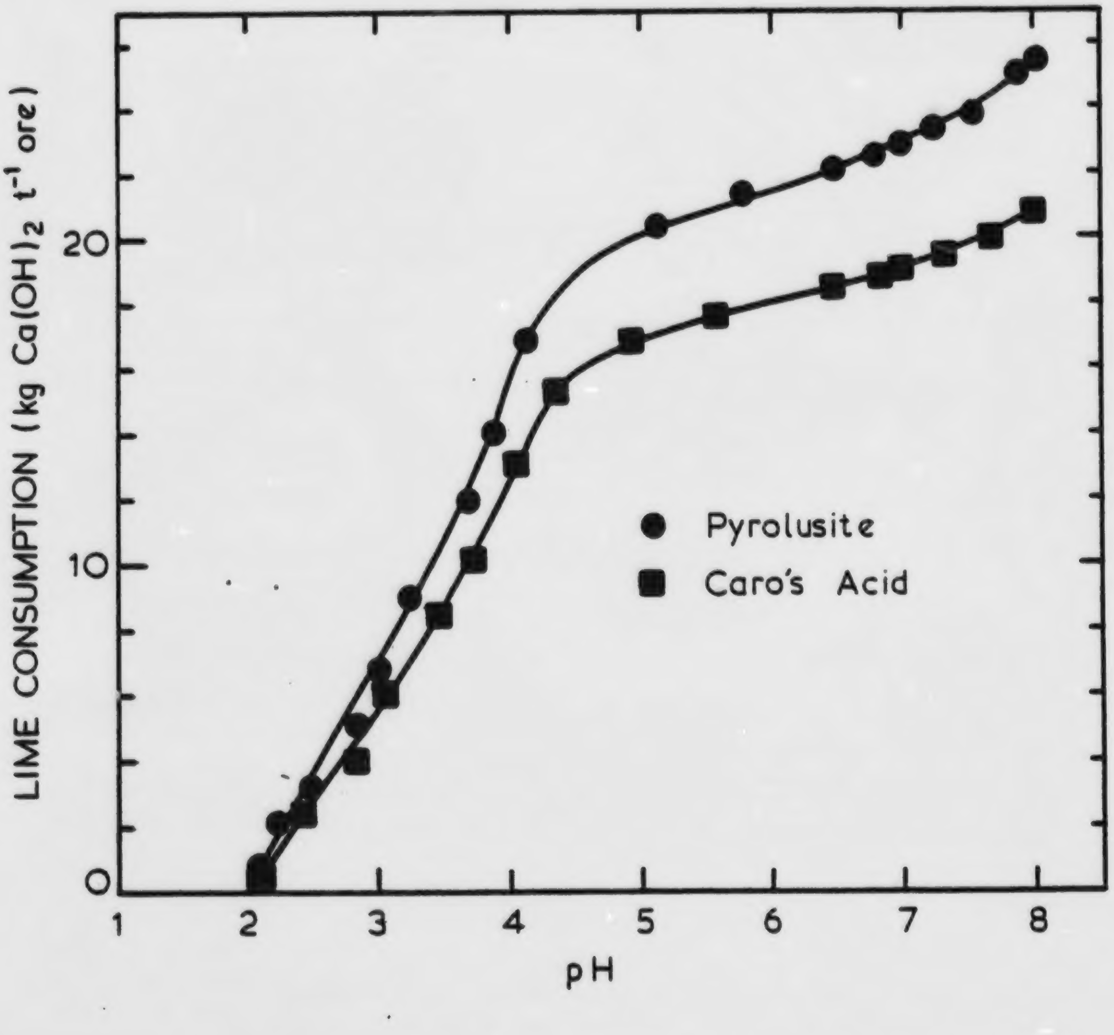

Figure 3. Neutralisation curves for Koongarra leach liquor/tailings slurries

5.5 Compositions of Neutralised Liquors

For Caro's acid as oxidant, Table 7 indicates that the manganese concentrations in tailings liquors ranged from 0.3-6 mg L^{-1} compared with 150-360 mg L^{-1} for pyrolusite. If Caro's acid were used as oxidant, neutralising to pH 8 may not be necessary as manganese concentrations at pH 7 were only 3-35 mg L^{-1}. Solubilities of other heavy metals do not change significantly in this pH range [2].

Apart from manganese, concentrations of sulphate and total dissolved solids (TDS) were lower in neutralised liquors where Caro's acid was the oxidant. Sulphate concentrations were reduced by 0.4-1.0 g L^{-1} and TDS by 20-25%.

TABLE 7

CONCENTRATIONS OF MANGANESE IN TAILINGS SLURRIES

(pH 8.0)

Oxidant	Manganese Concentrations (mg L^{-1})			
	Ranger	Nabarlek	Koongarra	Jabiluka
Pyrolusite	180-200	120-150	200	180-360
Caro's Acid	0.3-0.6	0.3-6.0	4.5	0.7

6. CONTINUOUS TESTWORK

6.1 Equipment and Procedures

The continuous leaching circuit consisted of four baffled tanks (4 litres per stage) connected in series, with gravity flow between stages and mechanical agitation in each stage. An ore-water slurry was pumped to the circuit from a storage tank at a rate of 640 mL h^{-1}, to give an average total residence time of 24 h. The temperature of the four tanks was controlled to within ±0.5°C by a water bath.

Continuous leaching tests of 120 h duration were carried out using pyrolusite and 25 wt % Caro's acid as oxidants with both Ranger and Nabarlek ores. The standard control conditions for each ore are summarised in Table 8.

The rate of leaching was determined by analysis of leach residues taken from each tank every 8 h. A daily composite leach liquor sample for metal ion analysis was prepared for each tank by combining samples taken at 8 h intervals. Batch neutralisation tests were carried out on slurries collected from the overflow of the final leaching tank.

6.2 Results

Figure 4 shows that uranium extractions remained constant for the duration of the 120 h experiments. The average cumulative extractions based on

TABLE 8

OPERATING CONDITIONS - CONTINUOUS TESTWORK

NABARLEK ORE (40°C, 45 wt % slurry)

Tank No.	Pyrolusite Run		Caro's Acid Run	
	pH	Redox Potential (mV)	pH	Redox Potential (mV)
1	1.80	420	1.80	420
2	1.80	420	1.80	420
3	N.C.	N.C.	N.C.	N.C.
4	1.80	N.C.	1.80	N.C.

RANGER ORE (45°C, 55 wt % slurry)

Tank No.	Pyrolusite Run		Caro's Acid Run	
	pH	Redox Potential (mV)	pH	Redox Potential (mV)
1	2.00	480	2.00	480
2	2.00	480	2.00	480
3	2.00	450	2.00	450
4	2.00	N.C.	2.00	N.C.

N.C. Not Controlled

residue grades in the discharge from each tank are summarised in Table 9. As in the batch tests, uranium extraction was unaffected whether pyrolusite or Caro's acid was used as the oxidant. The small differences in Table 9 are not statistically significant. For both ores, final extractions were the same as obtained in batch testwork.

In batch testwork, a high proportion of reagents is consumed in the first half hour and any errors in pH or redox potential can significantly affect results. Comparative reagent consumptions in continuous experiments are a more reliable guide to oxidant usage. Figures 5 and 6 show that oxidant and acid consumptions were very steady for the duration of the experiments. For each ore, R_{OX} was 3.3 and S_{ACID}/P was 1.25. Both values are in very good agreement with those obtained in batch experiments. Batch neutralisation tests on tailings liquors produced in continuous testwork gave S_{LIME}/S_{ACID} ratios of 0.71, in good agreement with batch leaching tests.

TABLE 9

URANIUM EXTRACTION IN CONTINUOUS TESTWORK

Tank No.	Nabarlek		Ranger	
	Pyrolusite	Caro's Acid	Pyrolusite	Caro's Acid
1	91.6	91.1	86.8	85.7
2	97.0	97.0	89.8	89.4
3	97.6	97.6	91.2	91.0
4	97.8	97.8	91.6	91.5

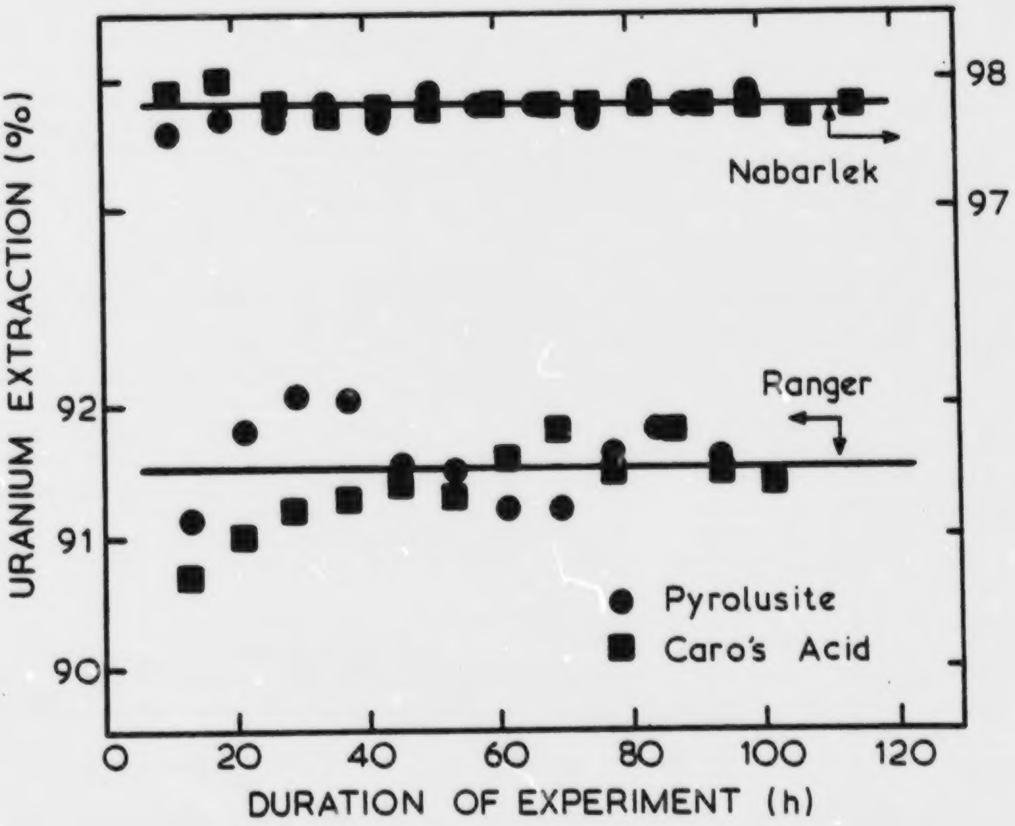

Figure 4. Extraction of uranium in continuous testwork

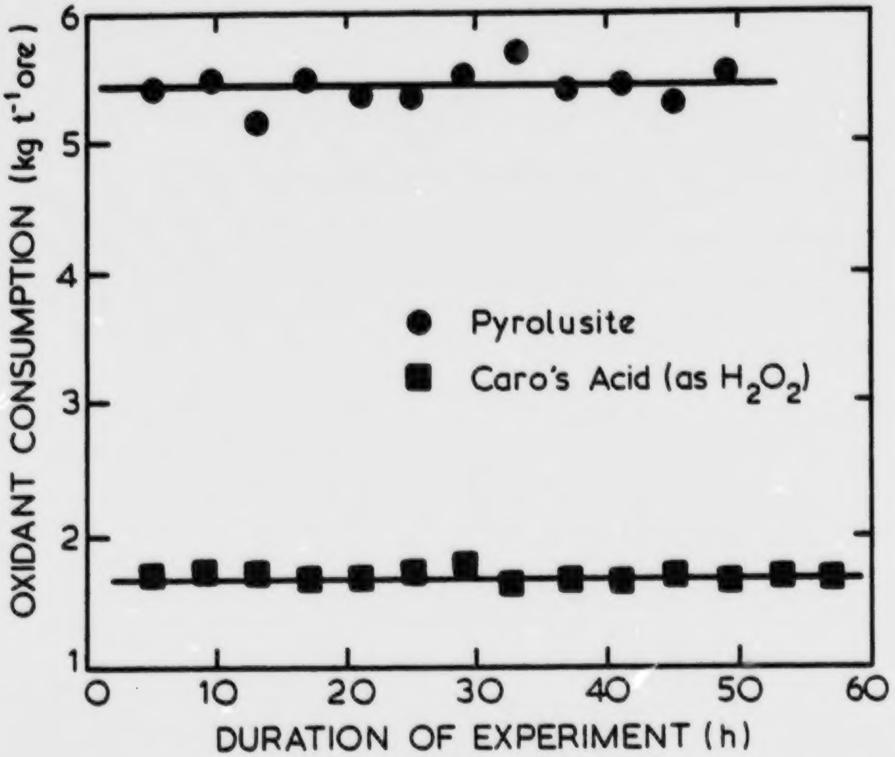

Figure 5. Oxidant consumption in continuous testwork - Nabarlek ore

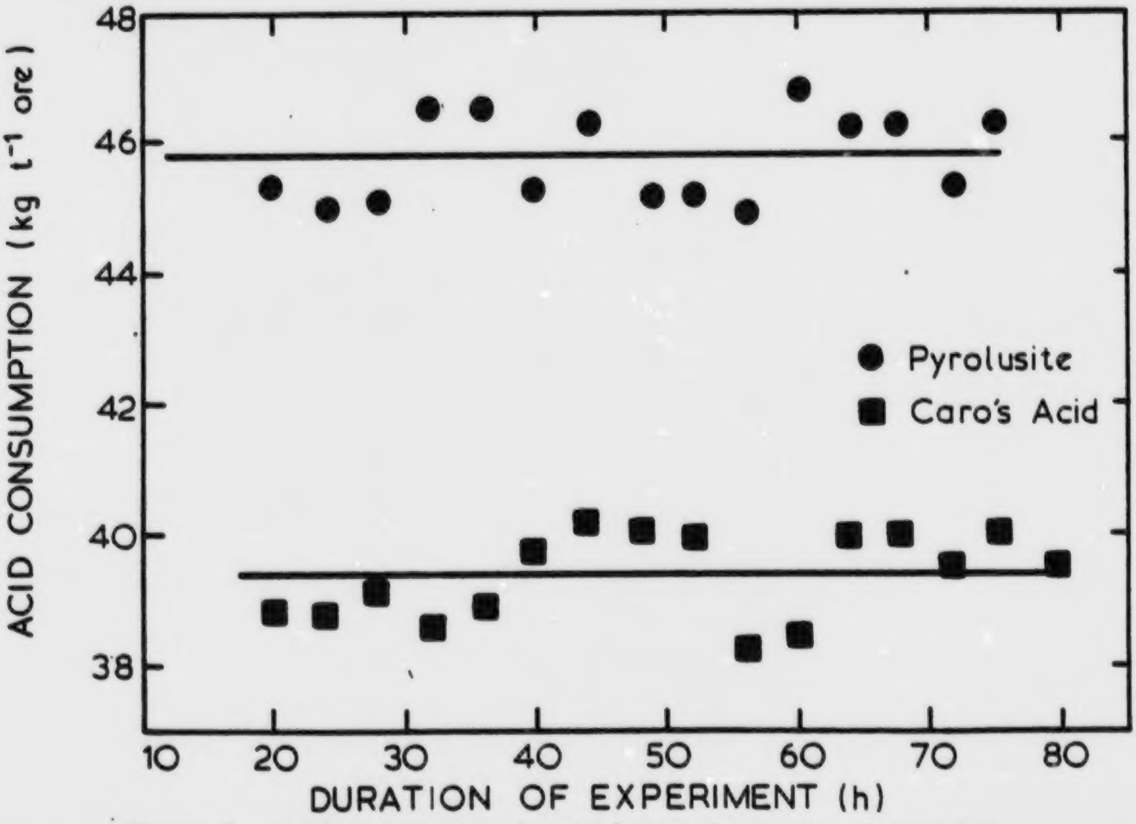

Figure 6. Acid consumption in continuous testwork - Ranger ore

7. PLANT TRIALS

7.1 Nabarlek Mill

The Nabarlek mill is located in the tropical Alligator Rivers region. The high grade orebody, averaging 2.34% U_3O_8, was completely mined by open cut methods and stockpiled before the commencement of milling in June 1980. Ore is treated by conventional acid leaching and solvent extraction processes. The leached solids are separated from the uranium-bearing solution by counter current decantation (CCD) washing with recycled raffinate. The tailings/raffinate slurry from the CCD circuit is neutralised with lime to pH 8.5 before discharge to the mined-out pit. Decant water from the pit is recycled to the plant via a storage pond.

The leaching circuit at Nabarlek consists of six mechanically-agitated tanks in series. Ore is ground to 50% less than 75 µm and leached as a 45 wt % slurry at 38-45°C without heating. Average ore throughput is 10 t h^{-1}. Sulphuric acid (98%) is added to the first two tanks to control pH at 1.6-1.8, and to the final tank for a pH of about 2.0. Originally a pyrolusite slurry was added to the first tank via a ring-main system to control redox potential at 450-475 mV. The pyrolusite was ground in a small ball mill and stored as a slurry in a mechanically agitated tank. About seven months after start-up this procedure was changed because of wear and blockage problems caused mainly by the relatively small slurry flow-rates required and difficulties in achieving satisfactory control of redox potential. Pyrolusite is now ground with the ore in the ball mill of the main grinding circuit. The ore/pyrolusite slurry then passes through a thickener and a 12 h storage tank before being pumped to the No.1 leach vessel. If required, further additions of pyrolusite are made using the original supply system.

Depending on ore grade and type, acid consumption ranges from 35-65 kg t^{-1} ore, pyrolusite from 5-15 kg t^{-1} ore, and lime ($Ca(OH)_2$), from 25-50 kg t^{-1} ore.

7.2 Caro's Acid Supply System

Some details of equipment designed for Caro's acid manufacture and reagent consumption data are proprietary information and are presented in general terms only. Caro's acid was made on site at a rate that matched the demand of the leaching circuit. The Caro's acid generator consisted essentially of a stainless steel reaction vessel and a refrigeration system. The storage capacity of the generating plant was equivalent to 2-4 h supply. Sulphuric acid and 50 wt % hydrogen peroxide, stored on site in portable 17.5 m^3 capacity tanks, were fed to the reaction vessel at rates corresponding to a mole ratio of about 3:1. The temperature of the 25 wt % Caro's acid produced was maintained at 10-20°C by the refrigeration system.

Caro's acid was pumped into Nos. 1 and 2 leach vessels via automatic control valves linked to the redox potential measuring system. Provision was also made to add Caro's acid to tanks 3 and 5 via rotameters and manual valves.

7.3 Results

Two trials each lasting about six weeks were carried out in June and August 1981. Reagent consumptions were compared with those in the four-week periods before and after each trial. During the five-month evaluation period, a wide range of ore grades (1.5-3% U_3O_8) and types were encountered.

Caro's acid consumption was measured daily from the change in liquid level in the hydrogen peroxide storage tanks. An instantaneous record of output flow from the generator was also available on a chart recorder or could be indirectly estimated from the control valve positions. The latter methods were particularly useful in providing an early indication of changes in ore type, which resulted in changes in oxidant demand.

Pyrolusite was fed into the ball mill circuit during the daily 12 h ore grinding campaign. However, only its weekly consumption could be estimated. Sulphuric acid consumption was measured every 8 h by a flow integrator on the supply line to the leaching circuit. Acid required for Caro's acid generation was taken from this line downstream of the integrator.

The variability of ore grade made it difficult to obtain direct comparisons of reagent consumptions for equivalent ores. Consequently, average results were compared over four-week periods. The major findings of the trials are summarised below :

(1) The Caro's acid generator was very reliable in operation. For the eleven-week trial period, oxidant supply was maintained without interruption.

(2) With Caro's acid, much better control of redox potential was achieved. Potentials in tanks 1 and 2 were readily maintained within ±5 mV of set points. Reasonable control was very difficult to achieve with pyrolusite; variations from set point were ±40 mV.

(3) The same uranium extractions were achieved from the use of pyrolusite and Caro's acid. Extractions tended to be less variable with Caro's acid because of the better control of redox potential.

(4) R_{OX} was 3.5 to 4.0 which is higher than the theoretical value of 3.2 because of the wastage of pyrolusite caused by poor control of redox potential and carry-over of unreacted pyrolusite.

(5) $\%S_{ACID}$ ranged from 12-21%. The wide range was attributable to large variations in oxidant demand. For the same ore type, oxidant consumption was roughly proportional to uranium grade. Absolute acid saving (S_{ACID}) was generally as predicted by equation 13b derived from laboratory testwork.

$$S_{ACID}/C = 4$$

(6) $\%S_{LIME}$ ranged from 5-23%. The variation in $\%S_{LIME}$ was even greater than that for $\%S_{ACID}$ because of

- the long delays in the plant before manganese was completely removed from process liquors, particularly as raffinate was recycled;
- the significant variation in acidity and flow of the raffinate fed to the neutralisation circuit;
- differences in the reactivity of batches of lime used in the trial period.

The major advantage of Caro's acid was the very good control of redox potential achieved in the mill circuit compared with pyrolusite. As pyrolusite is a solid, it reacts relatively slowly and, therefore, makes it difficult to maintain a constant potential for ores of varying oxidant requirements. This leads to either a redox potential above set point resulting in excessive oxidant and acid usage, or a low redox potential and consequently a slight loss in uranium extraction. In practice, this means that reagent savings for Caro's acid in the mill will be greater than those observed under carefully controlled laboratory conditions. Alternatively, uranium extraction could be slightly greater than for pyrolusite.

7.4 Advantages of Caro's Acid

The plant trials established that Caro's acid is a viable alternative oxidant to pyrolusite. From an operating viewpoint, Caro's acid was easier to handle, provided better control and a completely automated system, and ensured continual oxidant supply. It is clearly a superior system to pyrolusite and has the advantage of almost eliminating manganese from waste effluents.

The economics of Caro's acid versus pyrolusite is site specific and depends on the local cost of reagents. For Alligator Rivers ores, Caro's acid is more economic, especially if incorporated into a grassroots plant. Compared to a pyrolusite system, power costs should be similar and maintenance costs are expected to be lower. In addition, the close control of redox potential achieved with a Caro's acid system may increase average uranium extraction over an extended period.

Koongarra Mines Pty Ltd has recently decided to use Caro's acid in the Koongarra mill which is now at the final design stage. Queensland Mines Ltd is also considering changing to Caro's acid at Nabarlek.

8. CONCLUSIONS

The most important results from the experimental studies and plant trials were as follows :

- Uranium extraction was unaffected whether pyrolusite or Caro's acid was used as oxidant.

- The following reagent savings were established in the laboratory for the use of Caro's acid as an alternative oxidant to pyrolusite in the leaching of Alligator Rivers ores.

  Oxidant ratio, R_{OX} = 3.2

  S_{ACID}/P = 1.25

  S_{LIME}/S_{ACID} = 0.71

- For Caro's acid as oxidant, the manganese concentrations in tailings liquors neutralised to pH 8 ranged from 0.3-6 mg L^{-1} compared with 150-360 mg L^{-1} for pyrolusite.

- Plant trials confirmed the reagent savings obtained in laboratory testwork and demonstrated that Caro's acid has a number of operational advantages over a pyrolusite system. The most important was a much closer control of redox potential which could improve average uranium extraction over an extended period.

9. ACKNOWLEDGEMENTS

The authors wish to thank Mr F.J. Gee (Manager - Nabarlek) of Queensland Mines Limited and his staff for their involvement and assistance during the mill trials. We also wish to thank Queensland Mines Limited for their permission to publish the results of the trial.

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DISCUSSION OF PAPER BY R. RING

J. Goode	Are Narbarlek still operating the demonstration plant?
R. Ring	No, this has been taken away, and they are using pyroluscite. A small mixer is used for mixing-in the Caro's acid.
L. Melis	Did the CCD circuit operate any better with the use of Caro's acid?
R. Ring	Yes, there seemed to be an improvement.
C. Lendrum	Was there any work done on ores containing pyrrhotite? At Agnew Lake, a number of years ago, H_2O_2 was used in leaching, but polythionates were produced.
R. Ring	No, we have had no experience with that type of mineralogy.
V. Lakshmanan	Was there any difference in the yellowcake quality with the use of Caro's acid?
R. Ring	There was no difference in the quality of the product. However, better control was achieved in the monitoring and in EMF control when using Caro's acid.
L. Bates	The ore and pyroluscite are now ground together. Is this an improvement?
R. Ring	Yes, there is better oxidant utilization.
I. Itzkovitch	Is there a difference in the solvent loss encountered using Caro's acid?
R. Ring	No, we don't believe there is, but it was not investigated.

CIM METALLURGICAL SOCIETY

HYDROMETALLURGY SECTION

URANIUM '82

12TH ANNUAL HYDROMETALLURGICAL MEETING

TORONTO

AUGUST 31, 1982

PAPER 14.4

THE MIDWEST URANIUM PROJECT -

DEVELOPMENT OF THE MILLING PROCESS

L.A. Melis P.Eng.
K.S. Fraser
V.I. Lakshmanan

THE MIDWEST URANIUM PROJECT - DEVELOPMENT OF THE MILLING PROCESS

Abstract

This paper describes the development of a milling circuit for the complex Midwest orebody, a high-arsenic uranium ore typical of that found in northern Saskatchewan. The bench-scale testwork is described in general terms but most emphasis is on the pilot plant program carried out at Ontario Research Foundation. The complexity of the ore necessitated the investigation of many alternatives, not only for uranium extraction, but also for the treatment of the waste solutions. These are described. Pilot plant data are provided for the selected mill flowsheet. Also described are waste treatment circuits which have been established to produce effluents of acceptable quality. The applicability and sizing of equipment for the various circuits in the flowsheet are discussed. Pilot plant testing of sub-aerial tailings deposition is also described.

THE MIDWEST URANIUM PROJECT - DEVELOPMENT OF THE MILLING PROCESS

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THE MIDWEST URANIUM PROJECT - DEVELOPMENT OF THE MILLING PROCESS

1. INTRODUCTION

The Midwest uranium deposit is located in northern Saskatchewan, 700 km north of Saskatoon, near the eastern edge of the Athabasca Basin (see Figure 1.1).

Development work on the project is directed by Canada Wide Mines Ltd., a subsidiary of Esso Resources Canada Limited. Participants in the project are Esso Resources Canada Limited, Numac Oil and Gas Ltd., Bow Valley Industries Ltd., Mink Mining Corporation and Midwest Mining Corporation.

The deposit occurs at a depth of about 200 m at the unconformity between nearly flat-lying sandstone and older underlying gneisses. It consists of a series of flat lenses of massive ore overlain by a halo of lower-grade sandstone ore. The massive ore is formed of massive and colloform pitchblende and nickel arsenides, along with clay, and averages about 10% U and 9% Ni. The principal minerals are pitchblende (UO_2-UO_3), coffinite ($(U(SiO_4)_{1-x}(OH)_{4x})$), gersdorffite ($NiAsS$), niccolite ($NiAs$) and rammelsbergite ($NiAs_2$). The sandstone ore, a friable clay-rich sandstone with sooty pitchblende, averages about 0.4% U and 0.6% Ni. In places, the mineralization extends upward to form a further halo of sub-economic tonnage, termed special waste. This material averages 0.04% U.

The proposed plan is to recover the ore by open pit mining. All the ore would be stockpiled over a two-year period and reclaimed and trucked to the mill on a daily basis over a ten-year period. The special waste would be stockpiled separately. Other mining methods are being assessed.

Development of the milling process and waste treatment flowsheets has been in progress since 1978. The successful completion of the pilot plant program at Ontario Research Foundation (ORF) has resulted in establishing milling and waste treatment flowsheets for the Midwest ore.

The background work and the pilot plant program are detailed in this report and the selected process is described.

2. BACKGROUND WORK

2.1 General

A bench-scale development program for the Midwest ore was carried out during the period August 1978 to May 1981 in the laboratories of Lakefield Research of Canada Ltd. (Lakefield), (Lakefield, Ontario), Colorado School of Mines Research Institute (CSMRI), (Golden, Colorado) and to a lesser extent, Ontario Research Foundation (ORF), (Mississauga, Ontario) and Saskatchewan Research Council (SRC), (Saskatoon, Saskatchewan). Direction of the development studies was shared by Canada Wide Mines Ltd. and its metallurgical consultants, A.H. Ross and Associates. Input was received from Envirotech, Enviroclear, Interox America, and Knight and Piesold Ltd.

The development program was directed towards meeting the criteria of;

i) high uranium recovery
ii) process flexibility
iii) product purity
iv) use of conventional technology, and
v) environmental acceptability.

2.2 Leaching

The development of a suitable leaching process for the Midwest ore consumed the greater part of the overall process development program in the initial stages. The development program was required to contend with variations in ore grades as the assessment of the deposit changed. A number of metallurgical composites were made up from drill-core sample-rejects of nominal 10 mesh size. The typical analyses of the composites used are given in Table 2.1.

TABLE 2.1

COMPOSITE ORE ANALYSES (%)*

Composite	U	As	Ni	Fe
6a	1.03	2.00	3.00	1.9
6b	0.88	1.67	0.84	1.67
6c	0.68	1.30	0.68	1.43
8	2.18	4.82	2.83	2.17

A broad program of atmospheric acid-leaching with sodium chlorate oxidant was conducted on the lower grade ore with some limited investigations on the high-grade uranium ore. The leaching parameters assessed included acid addition, sodium chlorate consumption, temperature and leaching time. A summary of the leach conditions required for a 98.5% uranium extraction is provided in Table 2.2. The high-grade ore (composite 8) required a higher oxidant addition and a longer retention time.

TABLE 2.2

ATMOSPHERIC LEACH CONDITIONS

Composite	H_2SO_4 kg/t	$NaClO_3$ kg/t	Temperature °C	Time h
6a	120	9	60	2
	80	9	60	3
6c	120	9	60	2
	120	9	30	6
	80	9	60	2
	80	3	60	6
	40	9	60	6
	40	9	30	24
8	120	40	60	24

---

* unless noted all analyses are percent weight on a dry basis.

The use of hydrogen peroxide as an alternate oxidant to sodium chlorate for the lower grade uranium ore was investigated. Conditions for this leach program were defined intitially as 60°C, 6h, 80-100 kg/t H_2SO_4 and an emf maintained at -450 mv. Corresponding sodium chlorate tests were undertaken to provide a basis of comparison. The results are detailed in Table 2.3.

TABLE 2.3

H_2O_2, $NaClO_3$ LEACH COMPARISON

Composite		$NaClO_3$	H_2O_2 10%	H_2O_2* 10%	H_2O_2* 60%
6a	Time, h	1.5 3 6	1.5 3 6	1.5 3 6	1.5 3 6
	% U ext.	79.6 98.8 99.2	78.1 80.3 98.2	72.9 91.9 98.2	75.5 81.7 98.0
	Oxidant Consumption kg/t	8	14	21	21
6c	Time, h	1.5 3 6	1.5 3 6	1.5 3 6	1.5 3 6
	% U ext.	95.9 98.9 99.1	83.4 96.2 99.0	86.1 96.8 98.9	90.7 96.2 99.0
	Oxidant Consumption kg/t	5	8	7	12

* stabilized hydrogen peroxide reagent

The hydrogen peroxide requirements were found to be about 50% higher than that for sodium chlorate. This was likely due to a rapid H_2O_2 decomposition at the elevated leaching temperature. In addition, longer leaching time was required with H_2O_2 in order to obtain a 98.5% maximum extraction.

The investigation of MnO_2 as an oxidant with the high-grade ore indicated a 97% uranium extraction but H_2SO_4 and MnO_2 requirements of 160 kg/t and 68 kg/t respectively were excessive.

The use of a conventional air-pachuca leaching technique was tried in order to simulate atmospheric leaching with air-agitated tanks. Stainless steel, mechanically-agitated autoclaves were used for the application of low pressure air sparging. No additional oxidant was added. The range of leach conditions investigated are provided in Table 2.4.

TABLE 2.4

LEACH CONDITIONS FOR PACHUCA SIMULATIONS

Condition	Composite 6a, 6c	Composite 8
Temperature, °C	60	100
Acid addition, kg/t	40-80	160
Pressure, kPa	275-550	205
Air flow, L/s	1.9-2.9	1.9
Time, h	6-48	24-48
Pulp density, %solids	50	40

Leaching results of the lower grade ores indicated that a uranium extraction of at least 98.5% could be achieved by a pachuca style leach with a retention time of 24 hours.

The precipitation of arsenic was found to be a problem in these pachuca-simulation tests when an extended leach retention time was required. This led to liquid-solid (L/S) separation problems downstream. The results on the high-grade ore indicated acceptable uranium extraction but acid consumption was excessive.

An overall assessment of the above atmospheric leaching results led to the conclusion that excessive reagent additions would be required for high-grade ores and that arsenic precipitation would present a problem with extended leaching time. An assessment of pressure leaching with oxygen was then pursued as a leach alternative.

Autoclave pressure leaching tests were performed on the full range of uranium grades with leaching results on the low-grade ores indicating uranium extractions of 99%. The extractions appeared relatively independent of oxygen pressure within a pressure range of 140-550 kPa. Leaching results on the high-grade ore suggested that uranium extractions of 98.5% could be achieved with acid additions of 120 kg/t.

Subsequent to the completion of the leaching program at Lakefield further leaching work was carried out at CSMRI during the flowsheet development work.

The leaching optimization program at CSMRI included pressure leaching with oxygen, 2-stage leaching, and atmospheric leaching with Caro's acid (H_2SO_5).

The metallurgical composites provided for the leaching program are detailed in Table 2.5.

TABLE 2.5

ANALYSES OF CSMRI ORE COMPOSITES (%)

Composite	U	As	Ni	Co	Fe
9	2.13	3.97	2.14	0.28	2.25
14	2.19	4.57	2.35	0.41	2.43
18	2.07	3.03	1.65	-	3.65
19	2.07	4.41	2.58	-	2.83
21	1.68	1.72	1.21	-	3.12

The oxygen-pressure leaching studies were carried out on 0.5 kg, 1 kg and 5 kg batches. The range of operating parameters used in the testwork along with the optimum leaching conditions are detailed in Table 2.6.

TABLE 2.6

OXYGEN-PRESSURE LEACHING PARAMETERS

Parameters	Range	Optimum
Pulp density, % solids	40-55	50
Time, h	3-12	4
Temperature, °C	60-100	60
Pressure, kPa	450-725	450
Acid addition, kg/t	100-120	100
Oxygen consumption, kg/t	22-44	40

It was concluded that oxygen-pressure leaching under the conditions given would give the required 98.5% uranium extraction without arsenic precipitation problems and subsequent L/S separation problems.

Further tests on air-pachuca leaching at CSMRI confirmed the previous results obtained at Lakefield.

Oxygen-pressure leaching was therefore chosen as the preferred leaching process for Midwest ore. It was felt that air pachuca-leaching, with its longer retention time, would lead to:

i) increased attrition of leached ore
ii) high reagent consumption
iii) ferric arsenate precipitation
iv) poor heat utilization, and
v) poor oxygen utilization

To assess the possibility of 2-stage leaching, tests were done using a first-stage atmospheric leach followed by a second-stage oxygen-pressure leach. The leach parameters for these tests are detailed in Table 2.7.

TABLE 2.7

TWO-STAGE LEACH PARAMETERS

First Stage	Second Stage
30°C 2 h air, sparge agitated 19% solids	70°C 4 h mechanically agitated 100 kg/t H_2SO_4 480 kPa

Results indicated that the acid neutralizing capacity of the fresh ore was approximately 17 kg/t in a 2-stage leach process.

Since the precipitation of ferric arsenate did not present any problems with a one-stage oxygen-pressure leach and the acid neutralizing capacity of the ore was relatively low, it was decided that a 2-stage leach, in light of its complex nature, offered little benefit over the single-stage oxygen-pressure leach.

Following the completion of the above leaching testwork a final atmospheric leaching alternative was assessed in a series of tests performed using Caro's acid as the oxidant under atmospheric conditions. This work was directed by Canada Wide Mines Ltd. and supervised by Interox America.

Caro's acid was made up with 93% H_2SO_4 and 50% H_2O_2 yielding a Caro's acid solution of 26% H_2SO_5. The Caro's acid atmospheric leach successfully achieved a 98.5% uranium extraction in four hours under atmospheric pressure. These results justified further work to be done during the pilot plant program to provide an alternative to oxygen-pressure leaching.

2.3 Liquid-Solid Separation

Bench-scale thickening and filtration tests, on pulp samples of pressure-leached ore, were done at CSMRI. The purpose of the test work was to evaluate the following three L/S separation alternatives for the recovery of uranium pregnant solution from the leach pulp; i) counter-current decantation, ii) washing on drum filters, and iii) washing on horizontal-belt filters.

The standard leach conditions used to prepare leached-pulp samples for these tests were as follows:

i) Feed, % solids - 50
ii) Temperature, °C - 70
iii) Oxygen, kPa - 550
iv) Sulfuric acid, kg/t - 120
v) Time, h - 4

Polymer evaluations indicated Percol 351 as the flocculant for thickening and Percol 351 or 455 as the polymer for filtration.

Thickener evaluations included the assessment of both high-capacity and conventional thickeners by Envirotech representatives. Thickening tests were conducted for first, second, third and fourth stages of CCD washing. In addition information was gained on polymer dosage requirements for the entire circuit. The results are summarized in Table 2.8.

TABLE 2.8

THICKENER DATA SUMMARY

Data	Conventional	High-Capacity
Feed, % solids	8	15
Underflow % solids, design	45	45
Underflow % solids, expected	40	40
Polymer Type	Percol 351	Percol 351
Polymer Dosage, g/t		
1st stage	150-250	150
2nd stage	150	150
3rd + 4th stage	75	75
5th, 6th + 7th stage	15	15
Unit area, m^2/mtpd	0.45	0.35

Filtration tests provided filtration data and polymer dosage requirements for both drum filtration and horizontal-belt filtration. The results are given in Table 2.9.

TABLE 2.9

FILTRATION DATA SUMMARY

Data	Drum	Horizontal Belt
Filter media	POPR 929	POPR 929
Polymer type	Percol 351	Percol 351
Polymer dosage, g/t		
1st stage	650	350
2nd stage	200	100
3rd stage	100	50
Cake weight kg/m^2	5	10
Cake moisture, wt%	35	32
Total cycle time, min	6.6	2.94
Full-scale filtration rate, kg/m^2/h	45.4	200

Problems that were indicated for drum filtration included cake cracking, which would result in poor wash efficiencies, and highly variable pulp filtration characteristics which, at times, resulted in high polymer dosages. The general concern was that a drum filtration operation would be difficult.

Horizontal-belt filtration testwork indicated a lower polymer dosage requirement. This was attributed to the application of freshly flocculated pulp to the filter. The wash efficiency for horizontal-belt filtration was found to be lower than that for drum filtration.

Based on the above testwork, Table 2.10 details the required equipment and wash solution for a 99.9% uranium recovery in the washing circuit for the three L/S separation alternatives.

TABLE 2.10

WASHING CIRCUIT DATA

Data	CCD	Drum	Horizontal-Belt
Number of stages	8	3	3
Wash, kg/kg dry solids	4.167	5.42	5.4
Preg. kg/kg dry solids	3.667	5.88	5.93

Cost estimate comparisons for the three alternatives provided the information detailed in Table 2.11.

TABLE 2.11

COST COMPARISON - LIQUID/SOLID SEPARATION

Data	CCD (conv.)	Drum	Horizontal-Belt
No. of stages	8	3	3
Units per stage	1	3	1
Total units	8	9	3
Unit size, m^2	90	57	42
Dimensions	11m dia. conv.	3.66m dia.x 5.5m long	2m wide x 21m long
Comparative capital cost, $	4,011,000	5,456,000	4,700,000
Relative operating cost, $	348,000	712,000	552,000

Based on these cost comparisons and an assessment of other operating factors an 8-stage CCD circuit was recommended as the L/S separation step for the Midwest circuit.

2.4 Solvent Extraction

A small-scale continuous uranium solvent extraction pilot plant was set up in order to develop operating and analytical data for establishing design parameters.

The multiple-stage circuit was set up using uniformly sized mixer-settler cells of glass construction. The mixing compartments had an effective volume of 0.195 L whereas the settlers has an effective settling capacity of 0.5 L. Impellers were four-bladed glass turbine-type and were operated at the minimum speed necessary to obtain acceptable mixing as well as good interstage pumping. Four stages of extraction were included with a de-entrainer installed after the first stage of extraction and after the last stage of organic scrubbing.

Following loading of the organic in the extraction circuit, the organic was scrubbed in two stages using 15 g/L sulfuric acid and one stage using deionized water. The combined scrub solutions from these stages were returned to the extraction circuit to simulate plant practice.

Ammonium sulfate solution with concentrations of 150 g/L and 220 g/L were used as strip liquor in the uranium-stripping circuit. Initially four stages of stripping were employed but a fifth stage was found to be required in order to facilitate an effective means of molybdenum rejection in the stripping circuit. The stripping mixer emulsions were pH controlled with ammonium hydroxide from a pH of 3.2 in the first cell to pH 4.8 in the fifth cell.

A pregnant strip concentrate was precipitated with ammonium hydroxide or gaseous ammonia and the ammonium diuranate product was washed several times with deionized water.

The barren strip solution from precipitation was cooled to 20°C and the ammonium sulfate concentration was adjusted to the desired level by dilution with deionized water. This fresh barren strip solution was then returned to the circuit and used as barren strip liquor feed.

Regeneration (50% and 100%) of the stripped organic was a batch process using a 10% solution of sodium carbonate. The sodium carbonate regeneration solution effected molybdenum removal from the stripped organic to provide barren organic for the extraction circuit.

Crystallization of bleed samples of barren strip liquor was done in order to assess the practicality of ammonium sulfate by-product production for ammonia removal. The crystals of this fertilizer by-product were analyzed and contaminant levels were found to meet proposed government guidelines.

The circuit design for the solvent extraction mini-plant is given in Table 2.12.

TABLE 2.12

CIRCUIT DESIGN FOR SOLVENT EXTRACTION MINI PLANT

Stage	No. of Stages	Mixer Retention Time (min)	External O/A	Internal Recycle O/A
Extraction	4	2.0	1.0 - 1.5	-
Acid Scrub	2	4.0	4.5 - 7.5	1.2 - 2.0
Water Scrub	1	4.0	9.3 - 15.0	1.7 - 1.8
Stripping	5	4.0	4.7 - 12.5	1.1 - 1.7
Regeneration	1	2.0	10	-

Typical analytical results for the mini solvent extraction circuit with 100% regeneration of the stripped organic are given in Table 2.13.

TABLE 2.13

SOLVENT EXTRACTION ANALYTICAL RESULTS (g/L)

Stream	U	SO_4	As	SiO_2	Fe	Ni	Mo
Regenerated Organic	0.116	--	0.004	--	--	--	0.078
Loaded Organic	3.44	--	0.040	--	--	--	0.060
Barren Strip Solution	0.001	159.0	--	--	--	--	--
Solution from Precipitation	--	--	--	--	--	--	--
Strip Concentrate	13.4	173.0	0.006	--	--	--	0.022
Water Scrub Effluent	--	--	--	--	--	--	--
Dilute H_2SO_4 Scrub	--	--	--	--	--	--	--
Acid Scrub Effluent	0.048	8.35	0.254	--	--	--	--
Pregnant Solution	3.18	45.2	3.08	0.139	1.52	2.25	0.057
Composite Raffinate	0.0003	40.7	2.84	--	1.39	2.00	0.004
Na_2CO_3 Solution	4.81	--	0.067	--	--	--	0.577

Typical analytical results obtained for the yellowcake precipitated from the pregnant strip liquor are given in Table 2.14.

TABLE 2.14

YELLOWCAKE ANALYSES FROM MINI-PLANT (%)

U	As	Mo	SO_4	Zr	Ti	P	V
79.5	0.099	0.10	1.25	0.003	<0.01	0.03	<0.01

2.5 Waste Management

Waste treatment circuits were tested, along with a bench-scale tailings storage facility during the course of the leaching and solvent extraction testwork at CSMRI.

Due to the presence of large amounts of heavy metals, arsenic and radionuclides in the Midwest ore it was concluded that a conventional tailings disposal technique would not provide long term containment of contaminants in a tailings storage facility.

An alternative proposed by Knight and Piesold Ltd. was the sub-aerial layered-tailings deposition technique. This method could provide long-term containment of contaminants by leaving the deposited tailings in a partially saturated, consolidated condition with moisture retaining capacity.

Two layered-tailings models were constructed in order to test two extremes of final tailings pH values, pH 8 and pH 11.

These models included two experimental process-solution treatment circuits and provided representative seepage and runoff samples for chemical analysis and recycling to the process-solution treatment circuit. The flowsheet is shown in Figure 2.1.

The combined neutralized tailings (including precipitates from the solution treatment circuit) were deposited in layers in the tailings models to simulate the sub-aerial technique planned for the Midwest Project.

At the completion of the test program an assessment by Knight and Piesold Ltd. concluded that there were no major differences between the pH 8 and the pH 11 model. The pH 8 model indicated marginally greater densities and slightly lower permeability coefficients whereas the pH 11 material showed higher strength parameters and lower volumes of the combined seepage and runoff.

On the basis of these results bulk neutralization of tailings to pH 11 was selected as the final pH for tailings neutralization.

Following the tailings model program a detailed investigation was carried out into the secondary treatment of neutralized raffinate, and seepage and runoff, for the removal of arsenic and radium 226 to an acceptable level.

Radium was precipitated by the addition of barium chloride. Operating parameters of pH, time and Ba^{2+} content were tested.

Arsenic was precipitated by the addition of ferric iron as $FeCl_3$ or $Fe_2(SO_4)_3$. Operating parameters of pH, time and Fe/As ratio were investigated.

The range of operating parameters for the radium-arsenic precipitation circuit is provided in Table 2.15.

TABLE 2.15

Ra-As PRECIPITATION - TEST CONDITIONS

Ra Precipitation	As Precipitation
pH 8,11 time 0.5-2 h Ba^{2+} 100-400 ppm	pH 7-11 time 0.5-2 h Fe/As 2-24

The results of the testwork indicated radium precipitation was successful at pH 11 with a 0.5 h retention time and 200 ppm Ba^{2+}. Arsenic precipitation was more efficient at pH 9 with 1.5 h retention time and a Fe/As ratio in the range of 15-20. Treatment of seepage and runoff before or after the bulk neutralization step gave similar As and Ra results for the final effluent.

These results served as the basis for the treatment circuits investigated in the pilot plant.

In conjunction with this work, tests were accomplished for the treatment of simulated minewater made up by contacting water with low-grade and high-grade ore.

The range of operating conditions tested for the minewater treatment is given in Table 2.16.

TABLE 2.16

MINEWATER TREATMENT - TEST CONDITIONS

Neutralization	Ra Precipitation	As Precipitation
pH 8-11 time 1-2 h	pH 8-11 time 0.5-2 h Ba^{2+} 100-400ppm	pH 8-11 time 1-2 h Fe/As 1-6

The results were similar to the Ra/As precipitation testwork. These were used as the basis for the minewater treatment tests during the pilot plant program.

2.6. Special Waste

The potential recovery of uranium from the special waste (0.04% U) was investigated by carrying out heap leaching testwork.

A test program was initiated at Lakefield in order to assess its leaching characteristics. Bottle-leaching tests and several small-column leach tests were performed on drill-core sample-rejects representative of the special waste. The samples were nominally minus 0.64 cm in size and assayed <0.1% U.

Bottle leaching results, indicated a rapid extraction of up to 70% of the contained uranium in 72 hours of leaching time. Acid consumption during the leaching program ranged from 10-30 kg/t, depending on the leach solution pH.

Column leaching results confirmed the same rapid extraction rate found in the bottle leaching program. An overall uranium extraction of 76% within a period of 3 days of leaching was realized in a column leach performed in the pH range of 1.5-1.8.

Arsenic precipitation was not observed to be a problem and a satisfactory percolation rate of 7.5 L/h/m^2 was maintained.

After the successful completion of the bottle and column leaching programs a decision was then taken to go ahead with a pilot scale columns leach test in order to assess scale-up requirements. SRC was contracted to construct the leach column and carry out the leach test. Material from drill core was used. A screen analysis of the feed material is given in Table 2.17.

TABLE 2.17

HEAP LEACH FEED SCREEN ANALYSIS

Size Fraction, cm	Wt%
+5.1	32.2
+3.8	15.2
+2.5	14.1
+1.9	4.9
+1.3	3.2
+0.6	4.2
-0.6	26.2

The leaching was performed in a vertical PVC column 30.5 cm diam. by 245 cm high. During leaching the pH was maintained between 1.5-1.8 by the addition of sulfuric acid as required. Because of the available equipment the percolation rate was limited to 1.64 L/h/m^2.

The leach results indicated a uranium extraction of approximately 70%. This was in agreement with previous bottle and column leach tests. Due to the larger size of the material a leaching time of 60 days was required to achieve this extraction.

3. PILOT PLANT

3.1 General

The bench-scale test work was followed by a pilot plant program at ORF from August, 1981 to April, 1982. This program was supervised by Canada Wide Mines Ltd.

The pilot plant program was considered necessary in order to ensure reliable data for scale-up and plant design. The following items were considered important:

i) production of yellowcake to meet specifications.
ii) obtain final effluents of acceptable quality, both for mill waste solutions and mine water.
iii) ensure a good quality ammonium sulphate by-product.
iv) evaluate the sub-aerial tailings deposition on a pilot plant scale.
v) obtain data on the quantity and quality of seepage and runoff to be expected from the layered-tailings area.
vi) investigate the applicability of manufacturers' equipment for the various processes in the circuit.
vii) investigate the effect of solution recycle in the circuit.

For this program, representative samples of the orebody were obtained as 76 mm drill-cores during the winter and summer months of 1981.

The pilot plant program consisted of three separate campaigns:

i) No. 1 Campaign, run on 1.9% U ore.
ii) No. 2 Campaign, run on 2.1% U ore.
iii) Mine Water Treatment Campaign, run on simulated mine water.

No. 1 Campaign was used as a run-in campaign with Campaign No. 2 providing the design data.

A general process flowsheet for the pilot plant is shown in Figure 3.1. The main process stages included crushing, grinding, leaching, L/S separation, clarification, solvent extraction, uranium precipitation, ammonium sulfate by-product recovery, and waste management.

An assessment of equipment on pilot plant streams by manufacturers' representatives was considered an integral part of the pilot plant. Knight and Piesold Ltd. designed, monitored and reported on the tailings storage facility.

Investigation into by-product recovery for nickel, cobalt and molybdenum at various points in the circuit was not within the scope of this current program.

3.2 Equipment

The area of the pilot plant was divided into a process and waste treatment section (~12m x 12m) and a tailings storage area (~8m x 8m). In addition, separate areas were allocated for reagent make-up, sample preparation, and personnel change rooms.

The grinding circuit consisted of a 0.6 m diam. x 0.2 m Hardinge ball mill with a 32 kg charge of 3.8 cm balls in closed-circuit with a 700 micron, 0.3 m x 0.7 m DSM screen. A 3.8 cm Sala slurry pump transferred the mill discharge to the DSM screen. The mill feed rate was controlled by a Syntron vibrating feeder. A 0.9 m diam. neutral thickener dewatered the pulp prior to feeding to the leach circuit. Neutral pulp was transferred to the leach circuit batchwise with a 3.8 cm Sala pump.

Batch pressure-leaching was accomplished in a 210 L Pfaudler glass-lined autoclave with automatic temperature control. Oxygen was added from a gas bottle through a sparge tube in the autoclave and the sulfuric acid was added manually. Mechanical agitation was provided by either a Pro-Chem dual max-flow impeller of 904-L construction or a flat turbine/45° double-impeller of titanium construction. The off-gas was continuously monitored with an O_2/CO_2 analyser.

Atmospheric leaching was accomplished with the same vessel but at ambient pressure. The flat turbine/45° double-impeller was used for agitation. Oxidizing reagents were added via metering pumps.

Pulp transfer from the leach vessel to a leached-pulp storage tank was accomplished with a 3.8 cm FRP-coated Sala slurry pump. The leached pulp was stored in 22 L plastic containers as feed to the L/S separation circuit. The grinding and leaching circuits are detailed in Figure 3.2.

The remaining flowsheet was operated in a continuous mode with the leached pulp transferred manually to a 60 L agitated feed tank equipped with a pneumatic, time-controlled discharge valve to control the feed rate to the CCD circuit.

The CCD circuit was made up of 5-0.9 m diam. rubber-lined Dorr thickeners with 5-3.8 cm FRP-coated Sala pumps for underflow and overflow transfer. Underflow discharge was controlled with a pneumatic pinch valve. Flocculant was added through Clarkson reagent feeders from a common head tank to a 4 L flash-mix vessel that discharged to the feed well of the thickener. Acid wash solution was made up in a 1,000 L tank and added to No. 5 Thickener via a Cole-Parmer metering pump. The CCD flowsheet is given in Figure 3.3.

The pregnant solution from No. 1 CCD thickener overflowed into a 20 L agitated tank where Polyox solution was added from a drip bottle. The pregnant solution was transferred via a Peri-pump to a 2 L flash-mix vessel, flocculant added, and the solution then discharged to the feed well of a 0.6 m diam. FRP-coated clarifying thickener.

The clarifier overflow was then pumped, via a Hills McCanna metering pump, to a 15 cm diam. x 150 cm single-media sand filter (0.81 to 1.5 mm sand) and/or a U.S. precoat filter (0.23 m^2 Alsop filter with decalite and Alpha-floc medium). The clarified pregnant solution was then stored in a 450 L surge tank for feed to the solvent extraction circuit. The clarification circuit is shown in Figure 3.4.

The solvent extraction circuit included four extraction mixer-settlers, two acid-scrub and one water-scrub mixer-settlers, five stripping mixer-settlers, and one regeneration mixer-settler. Two after-settlers were also installed for the raffinate. In addition, storage tanks and metering pumps were included for:

i) Barren organic
ii) Acid-scrub solution
iii) Water-scrub solution
iv) Pregnant strip liquor
v) Barren strip liquor
vi) Sodium carbonate regeneration solution
vii) Raffinate
viii) Ammonium hydroxide (by drip bottles)

The settler and mixer areas for the various stages were:

i) Extraction - settler 0.085 m^2 - (PVC construction)
- mixer 0.016 m^2
ii) Scrubbing, stripping, regeneration
- settler 0.057 m^2 - (FRP construction)
- mixer 0.04 m^2

Detailed flowsheets of the solvent extraction circuits are provided in Figures 3.5, 3.6 and 3.7.

The yellowcake precipitation circuit (Figure 3.8) consisted of a 450 L pregnant strip liquor holding tank, a metering pump, two 20 L precipitation tanks in series, a 2 L flash-mix vessel where flocculant was added, and a 0.6 m diam. thickener. The addition of ammonia diluted with air was metered through Matheson flowmeters. Washing and filtration of the yellowcake was carried out in the laboratory.

The barren strip liquor was recycled to the solvent extraction circuit through a 7.5 cm diam. dual media filter containing sand and anthracite. A periodic bleed-sample was taken for ammonium sulfate crystallization by evaporation in a laboratory oven.

The raffinate treatment circuit (Figure 3.9) consisted of 3-40 L bulk neutralization tanks with mechanical agitation and air sparging. Lime was added through a peristaltic pump and a Clarkson feeder from a 100 L head tank. The pH was controlled with a Radiometer pH meter and titrator. The neutralized slurry was filtered through a 0.9 m diam. x 0.6 m Dorr-Oliver drum filter to produce a filter cake. The gypsum cake was then added manually to the tailings neutralization tanks. The filtrate was pumped to the Ra-As precipitation circuit for further treatment (Figure 3.10).

In the Ra-As precipitation circuit the drum filter filtrate and seepage and runoff from the tailings area were treated together in a cascade series of 4 tanks with effective volumes equvialent to a two-hour retention time. Flash-mix vessels (2 L) were provided for the barium chloride addition and for the ferri-floc and lime addition. The pH was controlled automatically with a Radiometer unit. The precipitated slurry was then pumped to a flash-mix vessel (2 L) where flocculant was added and the slurry then discharged to the feed well of a 0.6 m diam. clarifying thickener.

The overflow from the clarifying thickener was pumped through a 7.6 cm diameter x 150 cm dual-media filter containing sand and anthracite. The filter discharge was collected in a 90 L tank and transferred to a 2,000 L holding tank where the pH was adjusted to 7-7.5 prior to disposal. The underflow from the clarifier was drawn periodically and added manually to the tailings neutralization tanks.

The tailings neutralization circuit (Figure 3.11) consisted of 3-20 L neutralization tanks with mechanical agitation and air sparging. The combined CCD tails, gypsum cake and Ra/As precipitate were neutralized with the continuous addition of lime slurry. The neutralized tailings were then discharged to the tailings storage area with a 3.8 cm Sala pump.

The tailings storage structure consisted of an approximate area of 40 m^2. It was enclosed in a sealed room which was ventilated through a 0.3 m diameter pipe by a 45 m^3/min exhaust fan. The floor of the tailings structure included an initial layer of sand sloped to the required grade, followed by polyethylene sheeting overlain with a sand filter blanket.

Drainage outlets for seepage and runoff were made of slotted PVC pipe (2.54 cm diam.), leading to a sealed outlet through the polyethylene liner and plywood wall. Water tightness was achieved by the use of polyethylene sheeting which covered all internal surfaces. The seepage and runoff was collected in holding tanks and pumped to the raffinate treatment circuit.

Tailings slurry transfer to the tailings area was achieved through a short 1.3 cm diam. PVC pipe with open ends. The use of 0.3 cm diam. holes in the pipe to simulate a spray bar was not successful.

The enclosed area was heated with a propane burner and the tailings surface heated with lamps in order to achieve the required evaporation rates. Air turbulence over the tailings area was provided by fans with the air flow directed towards the tailings surface. Evaporation pans were installed to measure evaporation rates.

The mine-water treatment circuit is shown in Figure 3.12. Simulated mine-water, stored in 2,000 L holding tanks, was metered to a series of 4 precipitation tanks in series with flash-mix vessels used for reagent addition. The addition of lime slurry, barium chloride and ferri-floc was controlled by Clarkson feeders. One test was run with a pre-clarification step using a U.S. precoat filter ahead of the precipitation circuit. The precipitate slurry was dewatered and clarified in the same circuit as used for the raffinate treatment final effluent.

In addition to the above pilot plant equipment, the following manufacturer equipment was made available for testing during the pilot plant campaign:

i) Sand filters - Himsley Engineering Ltd.
ii) U.S. pre-coat filters - Lee Chemicals Ltd.
iii) Stellar filter - Ecodyne Limited.
iv) Centrifuge - Bird Machine Co. of Canada Ltd.
v) Larox pressure filter - Outokumpu Equipment Canada Ltd.
vi) Krebs mixer-settler pilot unit - Exxon Minerals Co. with the assistance of Krebs.

Representatives from Envirotech Corporation, Enviro-Clear, Axel Johnson Inc. and Dorr-Oliver Canada Ltd. were present to assess the dewatering characteristics of various mill streams. Pfaudler Co. representatives were available for consultation with respect to the leach autoclave.

Flowrates for the Lurgi solvent extraction system were assessed by Lurgi in their laboratories in Frankfurt on pilot plant solutions.

3.3 Operations and Results

3.3.1 Ore Preparation

The pilot plant feed was drill-core of massive ore, sandstone ore, special waste and sandstone. These were combined to be representative of the average ore plus dilution. The dilution included waste rock which would be used as shielding against gamma radiation during ore mining.

All individual core samples were crushed to minus 6 mesh and assayed for U, As, Ni, Co and Fe. Selected samples were chosen and combined in one bulk sample. Homogenous blending of the ore was accomplished by a cone and quartering method. This provided the bulk sample for No. 1 Campaign which assayed 1.9% U. No. 2 Campaign bulk sample was prepared by adding more ore samples to increase the grade to 2.1% U.

Typical analyses of the expected mill feed is detailed in Table 3.1.

TABLE 3.1

TYPICAL MILL FEED ANALYSES

Element	%	Element	%
U	1.90	F	0.047
As	3.24	Mg	0.54
Fe	1.28	Na	0.12
Ni	1.65	K	1.21
Co	0.21	Cu	0.11
S	1.10	Mo	0.06
Ca	0.11	SiO_2	75.8
Zn	0.07	Al	3.19
V	0.078	Mn	0.04
Ti	0.28	Cr	0.02
P	0.03	L.O.I.	3.12
CO_2	0.5	Rb	0.045
Cl	0.03	Sr	0.012
Zr	0.046	T.Ra^{226}	7428 pCi/g

3.3.2 General Operation

The blended ore was fed to the grinding circuit via a vibrating feeder. The grinding circuit, a batch operation, was run intermittently in order to supply sufficient leach feed for a 24-hour operation. The typical grinding rate was 43.5 kg/day with the circuit being run an average of 10-12 hours per day.

Leaching was a batch operation where a total of 114 kg of ore per batch was leached in the 210 L Pfaudler autoclave. In No. 1 Campaign 29 batches of ore were treated for a total of 3,300 kg of ore. In No. 2 Campaign a total of 3,130 kg of ore were processed.

The CCD circuit was fed continuously with leached pulp at a rate of 0.33 L/min. The maximum continuous running time was 5 days during No. 1 Campaign and 8 days during No. 2 Campaign. The average sulfuric acid wash solution addition to No. 5 Thickener was 1.11 L/min.

No. 1 Thickener overflow (unclarified pregnant solution) averaged 1.02 L/min. No. 5 Thickener underflow was discharged to tailings neutralization at densities varying from 30-45% solids. The large variation in densities was due to blockage problems in the CCD underflow pipe.

Average flowrates for the various streams in the solvent extraction circuit were as follows:

i) Clarified pregnant solution - 1.01 L/min
ii) Raffinate - 1.11 L/min
iii) Acid + water scrub - 0.1 L/min
iv) Barren organic - 1.0 L/min
v) Barren strip - 0.27 L/min
vi) Sodium Carbonate - 0.1 L/min

The yellowcake precipitation circuit was run at an average pregnant strip liquor flowrate of 0.76 L/min. Yellowcake washing was carried out using bench filtration. A total of 52 kg of yellowcake were produced during No. 2 Campaign.

The raffinate neutralization circuit treated raffinate and sodium carbonate at an average flowrate of 1.11 L/min and 0.1 L/min respectively. Lime addition (as a 15% slurry) averaged 0.26 L/min. The gypsum cake density averaged 33% solids.

Feed to the Ra/As precipitation circuit included the drum filter filtrate as well as seepage and runoff from the tailings area at an average flowrate of 0.96 L/min. Sludge densities from the clarifying thickener were as high as 10% solids.

Feed to the tailings neutralization tanks included No. 5 Thickener underflow (CCD tails) at an average flowrate of 0.49 L/min and an average density of 35% solids, along with gypsum cake and Ra/As precipitate added intermittently. Lime addition (as a 15% slurry) average 0.11 L/min.

The total neutralized tailings slurry to the tailings storage area averaged 0.74 L/min. at 30% solids. Tailings deposition was by the layered-tailings method. Seepage and runoff were collected and returned to the raffinate treatment circuit for treatment.

3.3.3 Grinding

Typical operating data in the pilot plant grinding circuit were as follows:

i) Mill discharge density - 60% solids
ii) Screen undersize density - 23-32% solids
iii) Grind - 30% minus 200 mesh
iv) Neutral thickener underflow density - 53-55% solids

Work index values were obtained by ORF and Aerofall Mills Ltd. The values obtained were:

i) ORF (Bond Work Index)
   1.9% U ore - 18.3
   2.1% U ore - 19.9, 20.8
ii) Aerofall Mills Ltd. (Autogeneous Work Index)
   on selected core samples - 9.9
   sample containing a larger fraction of sandstone - 12.1

The neutral thickener unit area as determined by Dorr Oliver and Eimco representatives was 0.17 m^2/tpd for 50-55% solids at 0.03 kg/t Percol 351 flocculant addition.

3.3.4 Leaching

3.3.4.1 Pressure

Preliminary bench work was accomplished on 1.9% U material to confirm the required leach conditions for No. 1 Campaign ore. Optimization tests indicated the following leach conditions for obtaining 98.5-99% uranium extraction:

i) Particle size - 30% minus 200 mesh
ii) Density - 50% solids
iii) Temperature - 60°C
iv) Pressure - 450 kPa with oxygen
v) Time - 4 h
vi) Terminal acidity - 50 g/L
vii) Terminal emf - minus 400 to 430 mV
viii) Acid addition - 100 kg/t H_2SO_4
ix) Oxygen consumption - 15-20 kg/t

Under these conditions, extraction of As, Ni and Co were approximately 57%, 48% and 52% respectively.

A bench test was performed to determine the thermal characteristics of the pulp during leaching. The heat liberation in the initial stage of leaching was 69.2 kJ/kg of leach feed with a temperature differential of 27°C due to heat of dilution. Further heat input was required to achieve and maintain the desired leach temperature.

During No. 1 Campaign, problems were experienced with agitation and oxygen sparging in the 210 L Pfaudler autoclave with only a few leach batches resulting in the required extraction efficienty. This was attributed to low emf values caused by poor oxygen sparging.

In order to improve agitation and oxygen sparging, the autoclave impeller was replaced with a double-maxflow impeller designed by Pro-Chem Ltd. and two additional baffles were installed. The oxygen sparge tube was placed below the top impeller. This provided the required agitation as well as oxygen sparging and the required uranium extractions of 98.5-99% were achieved for pressure leaching in No. 2 Campaign. However, the improved leaching efficiency was accompanied with emulsion problems in the solvent extraction circuit. This was attributed to overleaching and attritioning of the pulp in the autoclave which caused an increase in silica levels in the pregnant solution. Pregnant solution as high as 0.8 g/L SiO_2 were noticed. In addition sizing analysis of leach feed and discharge showed a weight loss in the slimes fraction of the ore which confirmed that an overleaching problem existed.

Test coupons of various materials were positioned on the baffles and impeller blades in the autoclave. This was done in order to provide an indication of the most suitable materials-of-construction for the leach autoclaves. In that only a limited amount of ore was treated, it was accepted that these tests would only provide some indication as to the suitability of materials. The materials that stood up the best included titanium, Inconel 825, Avesta 254 SMO, Uddeholm 904 L and Inconel 625.

Non-metallic material tested included Triflex (3-ply) rubber, Urethane (bonded to steel) and Urethane-Irathane 141. Although the results could not be termed conclusive, the change in surface characteristics of these materials seems to indicate that they would not be suitable materials-of-construction for the autoclaves.

An assessment of ceramics by Pfaudler Co. representatives indicated that the use of ceramics for autoclave construction would be a viable alternative.

3.3.4.2 Atmospheric

In order to provide an alternative to the more complex pressure leaching operation, test work was done to arrive at an atmospheric leaching system that would provide the required uranium extraction at comparable costs.

Bench scale test work carried out at CSMRI in conjunction with Interox America indicated that atmospheric leaching with Caro's acid (H_2SO_5) as a combined lixiviant and oxidant was a possibility for Midwest ore (section 2.2).

Further test work at ORF substantiated these results and provided operating parameters for a pilot plant run during the second week of No. 2 Campaign. Atmospheric leaching was carried out in the 210 L autoclave open to atmosphere and using a double impeller. During this run, Interox America representatives were present to ensure the proper make-up of Caro's acid from stock 50% H_2O_2 solution and 93% H_2SO_4.

A hybrid leach (Caro's acid and sodium chlorate oxidant) is required in order to maintain acceptable acidity levels in the leach discharge as well as ensuring minimum acid requirements. In the leach, the Caro's acid is added in the initial stages and sodium chlorate is then added to maintain the emf above minus 400 mV during the remainder of the leach. Although a four-hour retention time was used, the actual leach was completed in two hours. Terminating the leach earlier than 4 hours could lead to lower reagent costs. It was determined that the chloride and sodium levels in the final effluent were within acceptable levels. There was no apparent downstream effects in the CCD circuit or the solvent extraction circuit.

Uranium extractions of 99% were achieved under the following conditions:

i) Temperature - 60°C
ii) Pulp density - 50% solids
iii) Acid requirement - 80 kg/t H_2SO_4
iv) Caro's acid - 29 kg/t
v) Sodium chlorate - 5 kg/t
vi) Time - 4 h
vii) Terminal emf - minus 400 mV
viii) Terminal acidity - 53 g/L

Average pregnant solution assays were 4.75 g/L U, 4.5 g/L As, 2.2 g/L Ni, 0.26 g/L Co, 1.5 g/L Fe, 0.11 g/L Mo, 0.52 g/L Al and 0.4 g/L SiO_2. These are similar to the pregnant solution assays obtained from pressure leaching, namely, 4.75 g/L U, 4.9 g/L As, 2.3 g/L Ni, 0.28 g/L Co, 1.8 g/L Fe, 0.11 g/L Mo, 0.60 g/L Al, 0.44 g/L SiO_2.

3.3.5 Liquid-Solid Separation

In No. 2 Campaign the uranium recovery in the 5-stage CCD circuit averaged 97.9% with a 5.3 P/O wash ratio. The average grade of No. 1 Thickener overflow was 4.77 g/L U, and the average grade of No. 5 Thickener underflow solution was 0.23 g/L U. Percol 351 was added to all thickeners with the total addition being 0.4 kg/t. As the thickeners were oversized for the feed rate, sliming problems were non-existent.

CCD data were obtained by manufacturer representatives during the pilot plant operation. These data are detailed in Table 3.2

TABLE 3.2

CCD DATA

Manufacturer	Thickener Unit Area (m^2/tpd)		U/F Density (% Solids)	Flocculant Dosage (kg/t)		
	Conventional	High Rate		No. 1	No. 2	Remainder
Envirotech	0.16	0.08	50	.05	.025	.015
Enviro-Clear	-	0.09	50-55	.075	.015	.015
Dorr Oliver	0.45	0.28	40-45	——	——	——
Axel Johnson	0.26*	——	60	.01	.01	.01

* Lamella thickener - equivalent to a loading rate of 16.3 L/min/m^2 for 15% solids feed pulp.

During No. 2 Campaign, a 50% raffinate recycle as wash solution was tested for a 24-hour period. Excessive frothing occurred in No. 5 and No. 4 thickener and frothing was beginning in No. 3 thickener. An increase in solution assays was also noticed in the pregnant solution with assays in the range of 6 g/L As, 3 g/L Ni and 0.4 g/L Co being obtained. Although raffinate recycle as wash remains as an alternative for plant operations, it is not recommended based on present data.

3.3.6 Clarification

The clarification of No. 1 CCD thickener overflow on a pilot plant scale proved to be a problem with total suspended solids (TSS) values being higher than normally expected.

The pregnant solution clarification circuit for No. 2 Campaign included pretreatment with Polyox WSR-301 (7 ppm) followed by Percol 351 addition (2.5 ppm) and initial clarification with a small clarifying thickener. The clarifier overflow was then passed through a sand filter followed by a pre-coat pressure filter. Attempts were made at keeping stock solution levels at a minimum to prevent re-precipitation of colloidal silica.

Average TSS values during No. 2 Campaign are given in Table 3.3. These are much higher than would be expected in a normal plant operation. Values obtained for the pre-coat pressure filter discharge during the pilot plant were only as low as 40 ppm TSS.

TABLE 3.3

PREGNANT SOLUTION CLARIFICATION (ppm TSS)

	Clarifier		Sand Filter	Pre-Coat Filter
	Feed	Discharge	Discharge	Discharge
Week #1 Pressure Leach	338	223	162	124
Week #2 Atmospheric Leach	247	135	103	99

Recommendations for pregnant solution clarification were obtained from manufacturers' representatives based on bench scale tests.

The recommended clarification circuit by Envirotech was a reactor-clarifier (40 L/min/m^2 rise rate) with 3 ppm Polyox dosage and 2 ppm Percol 351 dosage to provide initial clarification down to 80 ppm TSS. Although not tested, an Eimco SVG filter was recommended to further clarify the pregnant solution down to 5 ppm TSS for solvent extraction feed.

An alternative provided by Envirotech was pre-coat vacuum filtration using a drum filter. Filtration tests on pregnant liquor (650 ppm TSS) with Eagle-Pincher FW20 Celatone diatomite indicated clarities of 5 ppm TSS or less and form zone filtration rates of 100-150 L/min/m^2 which would represent full scale filtration rates of 35-55 L/min/m^2.

Test work by Lee Chemicals, agent for U.S. filters, indicated that a flowrate of 30 L/min/m^2 to a pressure pre-coat filter would provide clarities of <5 ppm. Body feeding with Decalite 436 at a rate of .025% of the feed rate would be required. This would be operated in series with a clarifying thickener. A cycle time of eight hours would be required for the filter.

Recommendations by Himsley Engineering Ltd. consisted of sand-filter clarification for the clarifying thickener overflow. Although not formally reported, TSS values of <5 ppm would be expected.

A brief series of test with a laboratory size Stellar filter (0.1 L/min) on pregnant solutions of 125 ppm TSS indicated clarities of 30-50 ppm with this unit. This unit was not assessed sufficiently to arrive at any recommendations.

3.3.7 Solvent Extraction

Previous test work had provided data on the operating parameters for the solvent extraction circuit. In the pilot plant various problems occurred which resulted in the solvent extraction circuit operating poorly in the first campaign. Improvements in the clarification circuit and an increase in settler area in the extraction cells resulted in a better operation in No. 2 Campaign. It became evident that temperature control at 30°C for the pregnant solution was critical in order to minimize emulsion problems in the extraction circuit. Table 3.4 details the effect of temperature on the collapse rate for a 12.7 cm emulsion band.

TABLE 3.4

PHASE DISENGAGEMENT - EFFECT OF TEMPERATURE

Temperature	O/A Ratio	Collapse Rate (O+A) ($m^3/m^2/h$)
23	1.2/1	5.9
30	1.5/1	7.7
35	1.5/1	11.2

The organic make-up was 5% Alamine 336, 5% Isodecanol and 90% Isopar M. A comparison of Kermac 470B kerosene and Isopar M kerosene indicated insignificant difference in the collapse rate between the two organic mixtures. Data for recycled or fresh organic were comparable.

The pilot plant operation of the solvent extraction circuit showed that 0.002 g/L U raffinate solutions could be obtained with four stages of extraction with an O/A ratio of 1.5:1. Removal of arsenic and other impurities was achieved with two acid scrub stages (15g/L H_2SO_4) followed by one water scrub stage. The three scrub stages had an O/A ratio of 2:1.

An effective uranium/molybdenum separation is obtained in a 5-stage pH controlled (pH 3.2 to 4.8) ammonium sulphate stripping circuit.

Regeneration of the stripped organic with Na_2CO_3 solution removed 90-95% of the contained molybdenum from the stripped organic. A raffinate after-settler was required in order to prevent organic contamination of the waste treatment circuit.

Typical solution assays for the solvent extraction circuit are given in Table 3.5.

TABLE 3.5

SOLVENT EXTRACTION CIRCUIT - TYPICAL ANALYSES (g/L)

STREAM	ANALYSES		
	U	As	Mo
Pregnant solution	4.75	4.5	0.1
Raffinate	.002	4	.01-.03
Loaded organic	4-5	.01-.07	.05-.15
Scrub solution	1	1	-
Scrubbed organic	4-5	.01-.07	-
Stripped organic	.01-.05	-	.05
Regenerated organic	.01	-	.002
Barren strip liquor	.005-.01	-	-
Pregnant strip liquor	15-20	.02-.05	.02-.05
Spent Na_2CO_3 solution	0.5-1	-	1.0

Crud problems were minimal when the solvent extraction circuit was operating properly. Crud analysis indicated the following make-up for the three stages:

i) Extraction -Si, As, Al, Ti, Zr
ii) Scrub - Si, Al, As
iii) Strip - U

Specific flowrates were obtained for three different solvent extraction systems (conventional, Krebs, and Lurgi) using pilot plant solutions with a solution temperature of 30°C.

The conventional specific flowrate was obtained at CSMRI, the Lurgi specific flowrate was provided by Lurgi, Frankfurt, from test work in their laboratories, and the Krebs specific flowrate was obtained at ORF on a Krebs mixer-settler test unit.

The results are summarized in Table 3.6.

TABLE 3.6

SPECIFIC FLOWRATES (versus total flow)

SYSTEM	$m^3/h/m^2$
Conventional	3.9
Lurgi	5.5
Krebs	7.7

With Midwest solutions the settler area requirements for a Krebs system were much higher than had been anticipated based on Krebs previous experience with other plants. As expected Lurgi data fell between a conventional system and the Krebs system.

3.3.8 Uranium Precipitation and Washing

Yellowcake precipitation in No. 1 Campaign had indicated potentially high Mo levels in the final product. A carbon column was therefore set up prior to No. 2 Campaign in order to assess molybdenum removal from the pregnant strip liquor by carbon adsorption.

In preparation for carbon adsorption, the pH of the solution was adjusted to 2.0 with sulfuric acid and then oxidized with H_2O_2. The solution was then passed through a 7.6 cm x 150 cm carbon column (Pittsburgh SGL 8 x 30) at a rate of 0.4 L/min prior to uranium precipitation with ammonia. Only 30 to 40% of the molybdenum was removed by this method.

The carbon column was then abandoned and the pregnant strip liquor precipitated directly with ammonia. It was found that acceptable Mo levels could be obtained with careful pH control in the stripping circuit as carried out in No. 2 Campaign.

The yellowcake precipitate was settled in a thickener using flocculant and washed counter-currently using Buchner funnels and a wash ratio of 6:1. The resulting sulfate levels were too high and re-washing was necessary. The required wash ratio was determined to be approximately 8:1. Typical yellowcake analyses are given in Table 3.7.

TABLE 3.7

YELLOWCAKE ANALYSES

Element	% U Basis	Element	% U Basis
U	71.7	Ca	0.01
As	0.14	B	0.003
Mo	0.17	Ti	0.01
S	0.18	Zr	<0.01
V	0.01	Si	<0.10
P	0.01	Mg	0.18
K + Na	0.32	Th	<0.01
CO_3	0.42	F	0.024
Fe	0.13	(Cl, Br, I)	<0.005

Qualitative results were obtained for dewatering and washing yellowcake with a pressure filter using a LAROX laboratory filter. These tests were done for general information as the unit was available during the pilot plant. Capacities of 10-15 kg/h/m^2 were obtained with 65-75% solids in the cake being achieved. The manufacturer felt that higher capacities would be possible if higher feed pressures were available.

The barren strip liquor (thickener overflow) was clarified in a dual-media filter containing sand and anthracite. This reduction of the TSS to <1 ppm resulted in a uranium content of <5 ppm U in the barren strip liquor. This would ensure acceptable uranium levels in an ammonium sulphate by-product from the crystallization of a bleed stream of barren strip solution.

Clarification of barren strip liquor using a pressure pre-coat filter was assessed by Lee Chemicals Co. The results indicated a flowrate of 25 L/min/m^2 with clarities of <5 ppm TSS being achieved. Body feeding was not necessary but should be allowed for in the design in case of upsets. A 2-day cycle time is estimated for this filter.

The use of a dual media sand filter, as assessed by Himsley Engineering, reduced the solids content in barren strip liquor down to <5 ppm TSS. The unit flowrate in this application would be 200 L/min/m^2.

Vacuum pre-coat filtration tests performed by Envirotech on a yellowcake thickener overflow containing 380 ppm TSS clarified the solution to <5 ppm TSS.

3.3.9 Waste Management

3.3.9.1 Tailings Neutralization

The CCD tails and precipitates from the raffinate treatment circuit were neutralized to pH 11 with lime in a series of 3 agitated tanks. Due to manual pH control the pH tended to be higher than 11. The neutralized tailings were discharged to the tailings storage area at an average of 30% solids.

With a manual addition of the raffinate precipitates and the surging nature of No. 5 thickener underflow, density control was difficult. Densities of the tailings pulp varied from 25% solids to 35% solids. At 35% solids the pulp was quite viscous and did not flow well. The optimum density range seemed to be 30-33% solids. The total neutralized tailings assayed approximately 0.1% U (solids) with the contained solution in the tailings slurry typically assaying 0.2 ppm U, 2-4 ppm As and 0.2 ppm Ni.

3.3.9.2 Raffinate Treatment

The raffinate treatment circuit, shown in Figure 3.9 and 3.10, consisted of two separate stages, a bulk neutralization stage followed by a Ra/As precipitation stage.

The mill solutions treated in this circuit along with their respective flows were:

i) Raffinate - 1.11 L/min
ii) Spent sodium carbonate solution - 0.1 L/min
iii) Seepage and run-off from the tailings area - 0.96 L/min

The combined raffinate and spent sodium carbonate solution was neutralised with lime (15% slurry) to pH 11 with a 1-hour retention time and the precipitated slurry filtered on a drum filter. The filter cake (at approximately 30% solids) was discharged to tailings and the filtrate was treated along with seepage and runoff from the tailings area for Ra/As removal.

The second stage precipitation procedure included the addition of $BaCl_2$ as a 5% solution followed by the addition of ferri-floc as a 7% slurry and lime to maintain the final pH at 9. A total retention time of two hours was required for this second stage.

The precipitated sludge was initially dewatered in a small thickener with the addition of Percol E-24 at 2 ppm. Flash-mixing of the flocculant with the slurry was necessary to form stable floccs. Sludge recycle was not necessary. In order to obtain good dewatering efficiencies and low TSS in the overflow it was critical for the sludge bed to be maintained in the clarifying thickener.

It was found that organic contamination in the raffinate treatment circuit would lead to upset conditions in the clarifying thickener.

Sludge underflow densities as high as 10-15% solids were achieved but an average density of 5% solids would be more practical due to the handling characteristics of the sludge.

Clarifier overflow was filtered through a dual-media sand filter made up of sand and anthracite. Excellent clarities were achieved with this unit. Occasional backwashing of the filter with an acid solution seemed to be required for efficient removal of the collected fine precipitates.

Typical analyses of solutions in the raffinate treatment circuit are detailed in Table 3.8.

TABLE 3.8

RAFFINATE TREATMENT - TYPICAL ANALYSES

SOLUTION	U ppm	As ppm	Ni ppm	Mo ppm	Ra pCi/g	TSS ppm	dry S.G.
Raffinate	2	4000	2200	30	500	-	-
Spent Na_2CO_3 solution	500-1000	-	-	1000	-	-	-
Seepage and runoff	<.5	1-2	<.5	-	10-20	-	-
Drum filtrate	2	5-10	0.4	-	40	250	-
Clarifier feed	-	-	-	-	-	1650	-
Clarifier overflow	-	-	-	-	-	20	-
Sand filter discharge	<.5	0.2	0.1	13	0.5	<1	-
Filter cake	-	-	-	-	60	-	2.4
Ra/As precipitate	-	-	-	-	125	-	3.0

The above results became available at the completion of No. 2 Campaign and, until then, a molybdenum problem in the final effluent (13 ppm Mo) had not been suspected. Previous bench test work at CSMRI had indicated acceptable Mo values in the effluent. The more critical elements were Ra and As, therefore these were the elements followed closely during the pilot plant.

Two reasons for the unsuspected Mo problem were:

i) the pilot plant ore assayed 0.06% Mo while previous bench work was done with ore assaying 0.02% Mo.

ii) the high Mo content in the spent sodium carbonate solution would lead to Mo removal problems.

At the end of No. 2 Campaign a series of bench scale tests were completed which indicated that acceptable molybdenum levels (<0.5 ppm) could be obtained with a highly oxidative step for both stages in the raffinate treatment circuit. However, these results were not consistent and a pilot plant trial, under these conditions, indicated Mo levels in the range of 5 ppm for the final effluent.

Carbon adsorption of the spent sodium carbonate solution indicated 90% removal of the molybdenum by passing the solution through a carbon column.

Further investigations provided information which led to a recommended modification to the raffinate treatment circuit for molybdenum removal to acceptable levels. Time constraints did not allow testing of this option but work by others(16) has shown acceptable results.

The revised circuit includes pre-treatment of the spent sodium carbonate solution through a carbon column followed by bulk neutralization of the combined raffinate, spent sodium carbonate solution, and seepage and runoff. This is followed by molybdenum precipitation with ferric sulfate, and then Ra/As precipitation followed by final effluent clarification.

The dewatering and clarification operations of the raffinate treatment circuit are considered to be as important as the chemical treatment parameters, for obtaining an environmentally acceptable final effluent.

Manufacturers' equipment was assessed in order to obtain comparative design data for the selection of the final process.

The raffinate neutralization step results in a slurry which requires dewatering for disposal of the filter cake (gypsum) to tailings. Filtration tests by Envirotech and Dorr-Oliver indicated filtration rates of 181 kg/h/m^2 and 16 kg/h/m2 at 24% and 33% solids respectively.

Thickening tests on the gypsum slurry were carried out by Enviro-Clear in a small laboratory unit. A feed loading rate of 61 L/min/m^2 was obtained with 2-3 ppm Percol E-24 addition. Underflow densities were in the range of 16-18% solids with overflow clarities of 10-20 ppm.

Thickening tests by Axel Johnson Inc. indicated a feed loading rate of 29 L/min /m^2 for a Lamella Gravity Settler/Thickener unit. Flocculant dosage was 5-10 ppm of Percol E-24 with overflow clarities of 10-20 ppm and underflow densities of 30% solids.

Dewatering of the gypsum slurry using a 15 cm Bird centrifuge was tested in No. 2 Campaign. The results indicated that a centrate clarity of 150 ppm TSS would be possible with a cake density of 30% solids from a feed containing 6% solids. Discussions with the manufacturer indicate that a centrifuge is an economical alternative to a thickening/filtration combination for dewatering gypsum slurry. In practice, a 95% solids recovery could be expected with centrate clarities of 1000 ppm or less.

Filtration tests on gypsum slurry using the LAROX filter press produced a gypsum cake containing 60% solids. A unit capacity of approximately 43 kg/h/m^2 was achieved with this unit.

Dewatering of the Ra/As precipitate was attempted by thickening, pressure filtration, vacuum filtration, and centrifuging.

Thickening tests are summarized in Table 3.9. Handling characteristics of this sludge indicate that a 5% solids underflow density would be recommended.

TABLE 3.9
Ra/As PRECIPITATE THICKENING TESTS

Manufacturer	Clarifier Unit Upflow (L/min /m^2)	Underflow % solids	Overflow TSS ppm	Flocculant Percol E-24
Envirotech	42	not given	10 - 20	2 ppm
Enviro-Clear	62	6.5-7	10	1.5-2 ppm
Dorr Oliver	28	not given	not given	not given
Axel Johnston (Lamella)	22	4-5	<20	2-3 ppm

Pressure filtration of the sludge was attempted with a LAROX pressure filter. These tests showed that this type of precipitate is difficult to dewater and dirty filtrates are produced. The data indicated unit capacities of 4 kg/h/m^2 with a 24% solids content in the cake. Vacuum filtration tests on the sludge were carried out by Envirotech. Drum filtration was ruled out and horizontal-belt filtration was found to give poor results. At best a unit filtration rate of 12 kg/h/m^2 was obtained with cake densities of 15% solids.

Centrifuge tests were carried out on the Ra/As sludge from the minewater treatment pilot plant. The results are given in section 3.3.9.3. Similar results could be expected for the Ra/As precipitate from the raffinate treatment circuit as the two precipitates are similar.

Final clarification of the pilot plant effluent was accomplished with dual-media sand filters in the pilot plant with clarities of <1 ppm TSS being achieved. A unit filtration rate of 247 L/min/m^2 was estimated by Himsley Engineering Ltd.

Tests by Lee Chemicals Ltd. indicated that a pressure pre-coat filter (U.S. Filter) would provide clarities of <5 ppm with a unit flowrate of 30 L/min/m^2.

Typical final effluent analyses are listed in Table 3.10.

TABLE 3.10

MILL FINAL EFFLUENT ANALYSES

Element	Assay	Element	Assay
TSS	<5 ppm	Cu	15 ppb
Cl	120 ppm	Fe	<50 ppb
SO_4	3450 ppm	Hg	<10 ppb
Mg	13 ppm	Mn	30 ppb
Na	680 ppm	Ni	45 ppb
Ca	720 ppm	Pb	60 ppb
K	55 ppm	Se	1 ppb
PO_4	<10 ppm	V	26 ppb
TOC	11 ppm	Zn	<15 ppb
Mo	13 ppm	Total U	<0.5 ppm
Al	<1 ppm		
As	170 ppb		
B	320 ppb	T.Pb^{210}	1.08 pCi/L (0.04 Bq/L)
Ba	45 ppb	T.Po^{210}	1.08 pCi/L (0.04 Bq/L)
Bi	<5 ppb	T.Ra^{226}	0.54 pCi/L (0.02 Bq/L)
Cd	<20 ppb	T.Th^{230}	<0.27 pCi/L (<0.01 Bq/L)
Co	<15 ppb		
Cr	16 ppb		

3.3.9.3 Mine Water Treatment

Simulated mine water was made up by contacting low-grade and high-grade ore samples with water using air agitation. The water used was made up by adding appropriate reagent-grade chemicals to deionized water to approximate Midwest ground water quality. Analyses of the mine water obtained are given in Table 3.11.

TABLE 3.11

SIMULATED MINEWATER ANALYSES (ppm)

ELEMENT	USING LOW-GRADE	USING HIGH-GRADE
As	2.45	80
Ni	16	40
Co	8.2	1.02
Fe	-	0.03
Cu	0.01	0.03
Pb	-	0.01
Zn	-	0.03
Mg	3.70	2.60
T Ra^{226} (pCi/L)	1639	1972

The treatment circuit included pH adjustment to pH 8 with lime followed by $BaCl_2$ addition and, after 0.5 h retention time, addition of ferri-floc as a 7% slurry, and lime to control the final pH at 9.

It was found that, prior to treating the mine water, a clarification step would be required for removing suspended solids. The high activity of high-grade fines in the mine water resulted in high Ra assays in the final effluent. Preclarification will remove high-activity solids prior to treatment and results in low Ra levels in the mine water final effluent.

Dewatering of the precipitate from the mine water treatment was carried out with a clarifying thickener followed by a dual-media filter for final clarification. When treating the low-grade mine water, underflow sludge recycle to the thickener feed was important due to the low solids content in the feed. The results of this clarification circuit were similar to the results obtained in the raffinate circuit. Feed densities varied from 1-5% solids depending on the mine water being treated.

Dewatering of the precipitate sludge was attempted with a 15 cm Bird centrifuge. Results from these tests are tabulated in Table 3.12.

TABLE 3.12

CENTRIFUGE TESTS

MINEWATER PRECIPITATE SLUDGE

Test	Data	Feed	Cake	Centrate
1	% Solids	9.3	11.7	3.1
	% Solids Distribution		90.8	9.2
2	% Solids	4.4	22.9	1.2
	% Solids Distribution		77.3	22.7
3	% Solids	4.4	28.4	1.3
	% Solids Distribution		74.3	25.7
4	% Solids	0.8	4.9	0.1
	% Solids Distribution		90.0	10.0

From these tests it was concluded that dewatering of mine water precipitate sludge or Ra/As precipitate sludge with a centrifuge was not practical. A sludge density of 5% solids was felt to be adequate in the context of the Midwest circuit and further dewatering would not be necessary. Final effluent assays from the mine water treatment circuit are detailed in Table 3.13.

TABLE 3.13

MINE WATER FINAL EFFLUENT ANALYSES

Element	Assay	Element	Assay
TSS	<5 ppm	Cd	<20 ppb
Cl	50 ppm	Co	<15 ppb
SO_4	1560 ppm	Cr	<30 ppb
Mg	4 ppm	Cu	<5 ppb
Na	7 ppm	Fe	<50 ppb
Ca	530 ppm	Hg	<10 ppb
K	3 ppm	Mn	15 ppb
Total PO_4	<10 ppm	Ni	<15 ppb
TOC	4.5 ppm	Pb	45 ppb
Al	<1 ppm	Se	<0.2 ppb
As	0.11 ppm	V	16 ppb
Mo	0.45 ppm	Zn	<15 ppb
B	50 ppb	Total U	<0.5 ppm
Ba	15 ppb	$T.Ra^{226}$	1.8 pCi/L (0.07 Bq/L)
Bi	<5 ppb		

3.3.9.4 Tailings Storage

Knight and Piesold Ltd., provided the design of the test tailings storage area and monitored the area throughout the pilot plant program.

The combined neutralized tails, gypsum filter cake and Ra/As precipitate sludge were deposited in the tailings storage area by a sub-aerial method to produce a laminated deposit of tailings. Successive layers were built up with seepage and runoff continually removed and returned to the raffinate treatment circuit. Typical contaminant content of the seepage and runoff is given in section 3.3.9.2. Rainfall was simulated by the addition of 500 L of water from a spray hose on a daily basis.

Pilot plant conditions resulted in an average tailings slurry feed of only 30% solids whereas 33% solids would have been the preferred density. The achieved evaporation rates in the test area were slightly lower than the required rate.

A water balance was calculated for the pilot plant tailings storage area using average flow measurements.

The water distribution as a percentage of total incoming water was calculated as follows:

i) evaporation, 21%; ii) runoff, 18%; iii) seepage, 21%; iv) storage, 40%

An ongoing program at ORF should provide additional information on the effect of time on the layered tailings. This program includes physical and chemical measurements of seepage and runoff from rainfall as well as monitoring any chemical change of the tailings material.

3.3.10 By-Product Recovery

To obtain samples of ammonium sulphate by-product, samples of clarified (dual-media sand filter) barren strip liquor were evaporated in a laboratory oven during the course of the pilot plant program. Analyses of the resulting crystallized ammonium sulphate by-product is provided in Table 3.14.

TABLE 3.14

AMMONIUM SULFATE CRYSTAL ANALYSES

Element	Assay	Element	Assay
U	21 ppm	Se	<0.5 ppm
As	<0.15 ppm	Zn	7 ppm
Ni	100 ppm	Ca	300 ppm
Co	38 ppm	Mg	100 ppm
Cd	<1 ppm	S	24.1 %
Cu	6 ppm	Total N	20.8 %
Cr	<1 ppm	TOC	<0.10 %
Hg	<0.01 ppm	T.Pb^{210}	<0.5 pCi/g
Mo	2 ppm	T.Po^{210}	0.81 pCi/g
Pb	1 ppm	T.Ra^{226}	0.27 pCi/g

4. SELECTED PROCESS DESCRIPTION

4.1 Flowsheets

An assessment of pilot plant data has resulted in the selection of the process and waste treatment flowsheets for the Midwest uranium project.

Figure 4.1 provides a detailed flow diagram for the selected milling process based on a mill feed assaying 1.6-2.2% U. The leaching flow diagram with the atmospheric leaching option using Caro's acid and sodium chlorate oxidants is depicted in Figure 4.2.

Figure 4.3 describes the waste treatment circuit for raffinate (including spent sodium carbonate solution, and seepage and runoff), CCD tailings neutralization and treatment of mine water. The recommended circuit modificiation for molybdenum removal, as discussed in section 3.3.9.2, is not included in this diagram but is described in section 4.3.

4.2 Milling Process

Reclaimed ore from the stockpile will be passed through a jaw crusher and fed to a two-stage grinding circuit consisting of an open-circuit semi-autogenous mill, followed by a secondary ball mill in closed-circuit with a 35 mesh Derrick or DSM screen. This will provide the required 30% minus 200 mesh grind.

Variations in ore hardness will necessitate intermittent running of the grinding circuit which will be designed for a 24-hour operation when treating hard ore. Neutral pulp storage and downstream blending will allow this flexibility in the grinding circuit.

Due to the viscosity of the pulp, the feed to the screen classifier will require dilution in order to obtain proper screening efficiency. Thickening of the dilute screen undersize pulp will be carried out in a neutral thickener to provide a leach feed density of 50-55% solids.

Data for both pressure and atmospheric leaching have been provided for the leaching circuit. Although both options will provide the required 98.5-99% uranium extration, atmospheric leaching, with a combination of Caro's acid and sodium chlorate as oxidizing agents, is preferred as it is an easier operation. A preliminary capital and operating cost estimate has shown similar economics for both options.

Liquid-solid separation and washing of the leach pulp will be carried out in a CCD circuit. Test work by equipment manufacturers has shown that vacuum filtration of the Midwest leach pulp is not viable due to its high clay content and low unit feed rates even with high flocculant dosage.

An eight stage CCD circuit with a P/O ratio of 4.75 and underflow densities of 50% solids will result in predicted soluble uranium losses of <0.3%. Interstage mixing will be required for better flocculation and more efficient washing.

A low mill tonnage throughput would favour the use of conventional thickeners in light of the favourable thickener unit areas determined during the pilot plant.

Recyclying of raffinate as wash solution is not recommended because of the high concentration of metals in the raffinate solution.

Clarification of the pregnant solution will require pre-treatment with Polyox for the precipitation of silicates. This will be followed by flocculant addition prior to inital clarification in a clarifying thickener such as an Envirotech reactor-clarifier. This unit would provide surge capacity for possible upsets in the CCD circuit.

Underflow from the reactor-clarifier can be returned to the head of the CCD circuit as required. The overflow from the reactor-clarifier will be passed through a pre-coat pressure filter using body feed of the filter aid. A U.S. filter type or a Durco Enzinger type can be used.

The use of sand filters remain as an alternative to the pre-coat filters for final clarification prior to solvent extraction.

The clarified pregnant solution temperature will be maintained at 30°C to ensure proper phase disengagement characteristics in the solvent extraction circuit.

The solvent extraction circuit will consist of four extraction stages, followed by two acid-scrub stages and one water-scrub stage. This will be followed by five stripping stages using ammonium sulfate strip solution and pH control of each stage with ammonia. This will provide an efficient uranium/molybdenum separation in the stripping circuit. Stripped organic will be regenerated with sodium carbonate solution in a single regeneration stage. The spent sodium carbonate solution will report to the raffinate after-settler prior to the waste treatment circuit.

Specific flowrates for three possible solvent extraction plants have been provided in section 3.3.7. In spite of the lower than expected specific flowrates for the Krebs' system, a preliminary cost comparison between the three systems seems to favour the Krebs' solvent extraction plant. A final assessment will be required during the detailed engineering phase.

The pregnant strip liquor will pass through an after-settler prior to uranium precipitation with ammonia at pH 7-7.5.

The yellowcake washing circuit will be made up of two thickeners and one centrifuge in series, with a wash water ratio of approximately 8:1. A multiple hearth drier, with calcining capabilities, will be used for drying the washed yellowcake product to a moisture of 2%.

In the first yellowcake thickener, flocculant addition will be required to ensure good clarity in the barren strip liquor. Final clarification of the barren strip liquor will be accomplished with a dual-media filter containing sand and anthracite to ensure low uranium values in the barren strip solution. The use of pre-coat pressure filtration is an alternative although the dual-media sand filter in the pilot plant provided excellent clarities (<0.5 ppm TSS).

Removal of ammonia from the solvent extraction and precipitation circuit will be carried out by the crystallization of a bleed stream from the barren strip liquor. Ammonium sulfate crystallization will produce a fertilizer product of marketable quality.

4.3 Waste Management

The mill waste solutions will be treated in the raffinate treatment circuit with waste precipitates being combined with the CCD tails for bulk neutralization with lime. The combined waste products will then be discharged to the tailings storage area.

Based on pilot plant results, the spent sodium carbonate solution will require pre-treatment with carbon adsorption for initial molybdenum removal (90%) prior to reporting to the raffinate treatment circuit.

The proposed circuit, which includes the recommended modification for molybdenum removal, consists of initial bulk neutralization with lime to pH 11 for the raffinate, spent sodium carbonate solution, and seepage and runoff from the tailings area. This step, which will remove the majority of the heavy metals, will be followed by molybdenum precipitation with ferric sulfate at pH 4.

Final removal of radium and arsenic will be accomplished with the addition of barium chloride, ferric sulfate and lime to control the final pH at 9.

Clarification of the final effluent will be done with a reactor-clarifier in series with a dual-media filter containing sand and anthracite.

The precipitate products will be transferred to the CCD tailings neutralization circuit where the combined pulp will be neutralized with lime to pH 11. The pulp, at 30-35% solids will be pumped to the tailings storage area.

Tailings disposal and storage will incorporate the sub-aerial technique of tailings deposition as proposed by Knight and Piesold Ltd. This technique provides a successive layered-tailings deposition which will produce a laminated deposit of tailings solids. The technique allows the free-draining liquid to be removed as seepage and the deposited tailings is left in a partially saturated, consolidated condition with moisture retaining capacity. Approximately 40% of the incoming solution contained in the tailings slurry and in the form of rainfall will report as seepage and runoff which is returned to the raffinate circuit for treatment. With time, as successive tailings layers are built up, seepage volume will decrease and runoff will correspondingly increase. At the completion of mining activities the tailings mass will have moisture retaining capacity and, with a proper seal cover, precipitation can report as uncontaminated runoff. Any minor seepage will be retained in the tailings mass.

5. ACKNOWLEDGEMENTS

The authors express their thanks to Canada Wide Mines Ltd. and the Midwest Project owners for permission to present this paper. The cooperation of Colorado School of Mines Research Institute and Lakefield Research of Canada Limited during the bench-scale program, and the cooperation of Ontario Research Foundation during the pilot plant, ensured the success of the program and this cooperation is acknowledged with thanks.

FIGURES

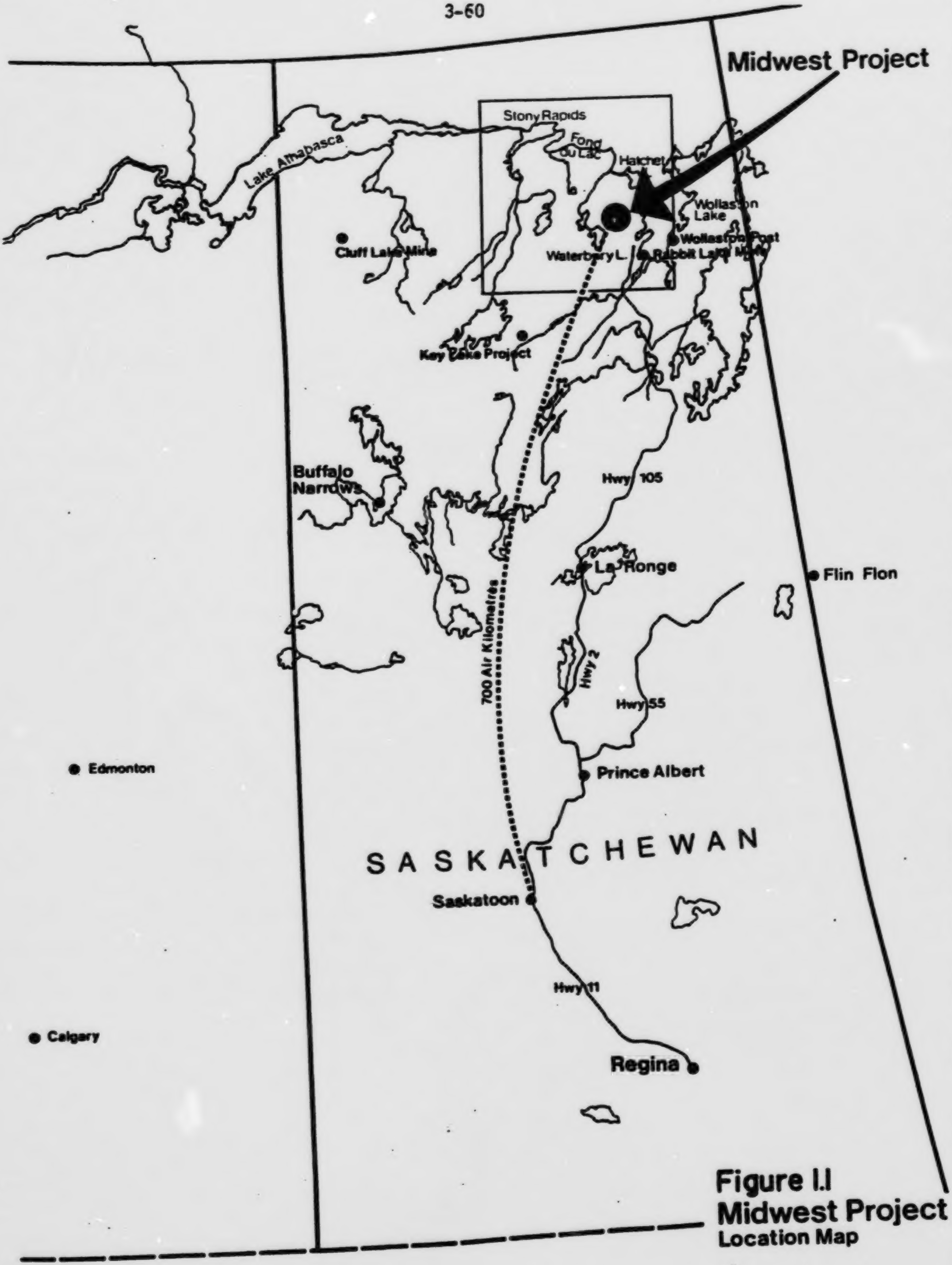

Midwest Project
Stony Rapids
Fond du Lac
Hatchet
Lake Athabasca
Wollaston Lake
Wollaston Post
Cluff Lake Mine
Waterbury L.
Rabbit Lake Mine
Key Lake Project
Buffalo Narrows
Hwy 105
La Ronge
Flin Flon
700 Air Kilometres
Hwy 2
Hwy 55
Edmonton
Prince Albert
SASKATCHEWAN
Saskatoon
Hwy 11
Calgary
Regina

Figure I.I
Midwest Project
Location Map

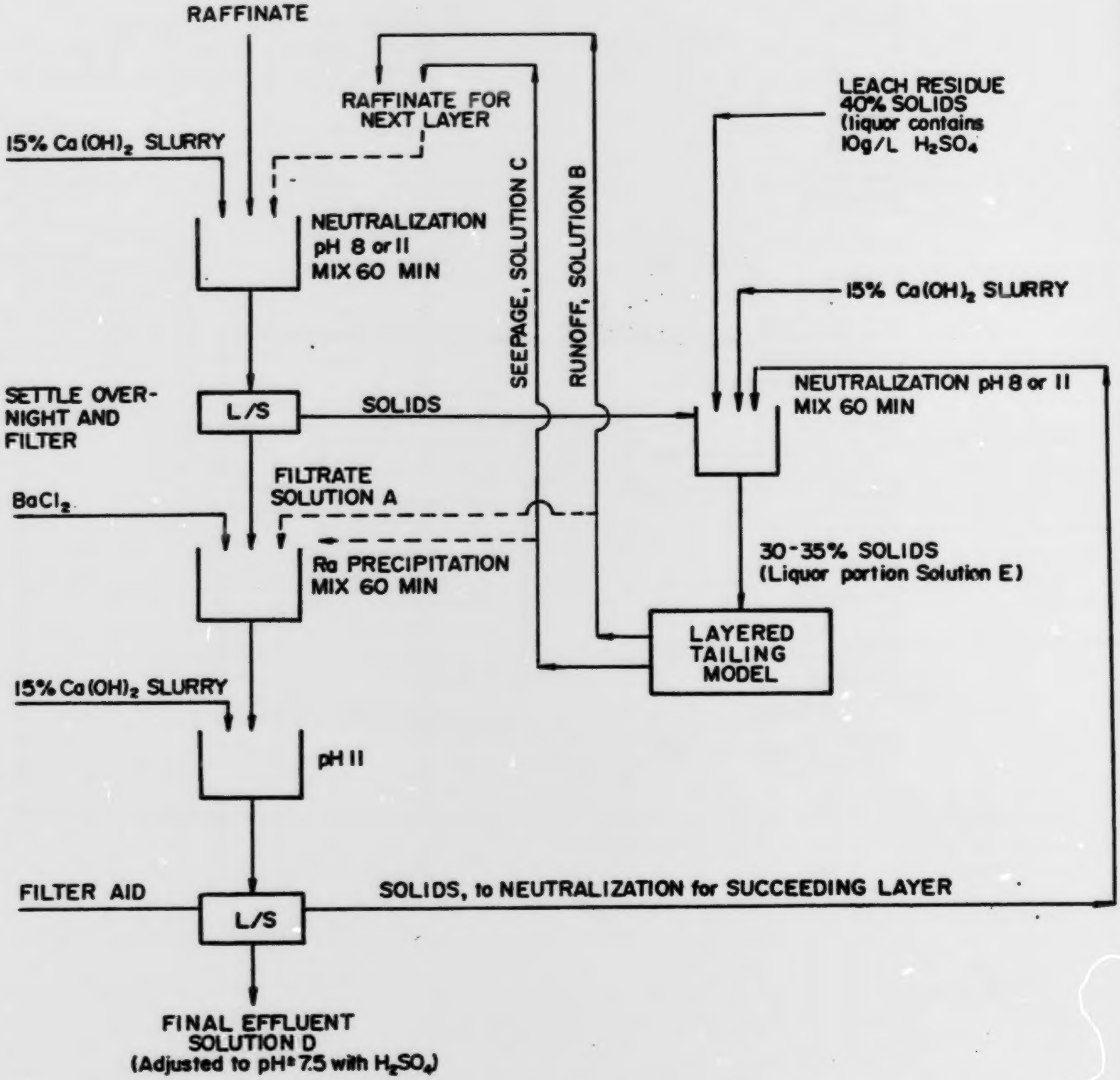

RAFFINATE
RAFFINATE FOR NEXT LAYER
15% $Ca(OH)_2$ SLURRY
NEUTRALIZATION pH 8 or 11 MIX 60 MIN
SEEPAGE, SOLUTION C
RUNOFF, SOLUTION B
LEACH RESIDUE 40% SOLIDS (liquor contains 10g/L H_2SO_4
15% $Ca(OH)_2$ SLURRY
NEUTRALIZATION pH 8 or 11 MIX 60 MIN
SETTLE OVER-NIGHT AND FILTER
L/S
SOLIDS
FILTRATE SOLUTION A
$BaCl_2$
Ra PRECIPITATION MIX 60 MIN
30-35% SOLIDS (Liquor portion Solution E)
LAYERED TAILING MODEL
15% $Ca(OH)_2$ SLURRY
pH 11
FILTER AID
L/S
SOLIDS, to NEUTRALIZATION for SUCCEEDING LAYER
FINAL EFFLUENT SOLUTION D (Adjusted to pH=7.5 with H_2SO_4)

Figure 2.1
Tailings Model
Testing

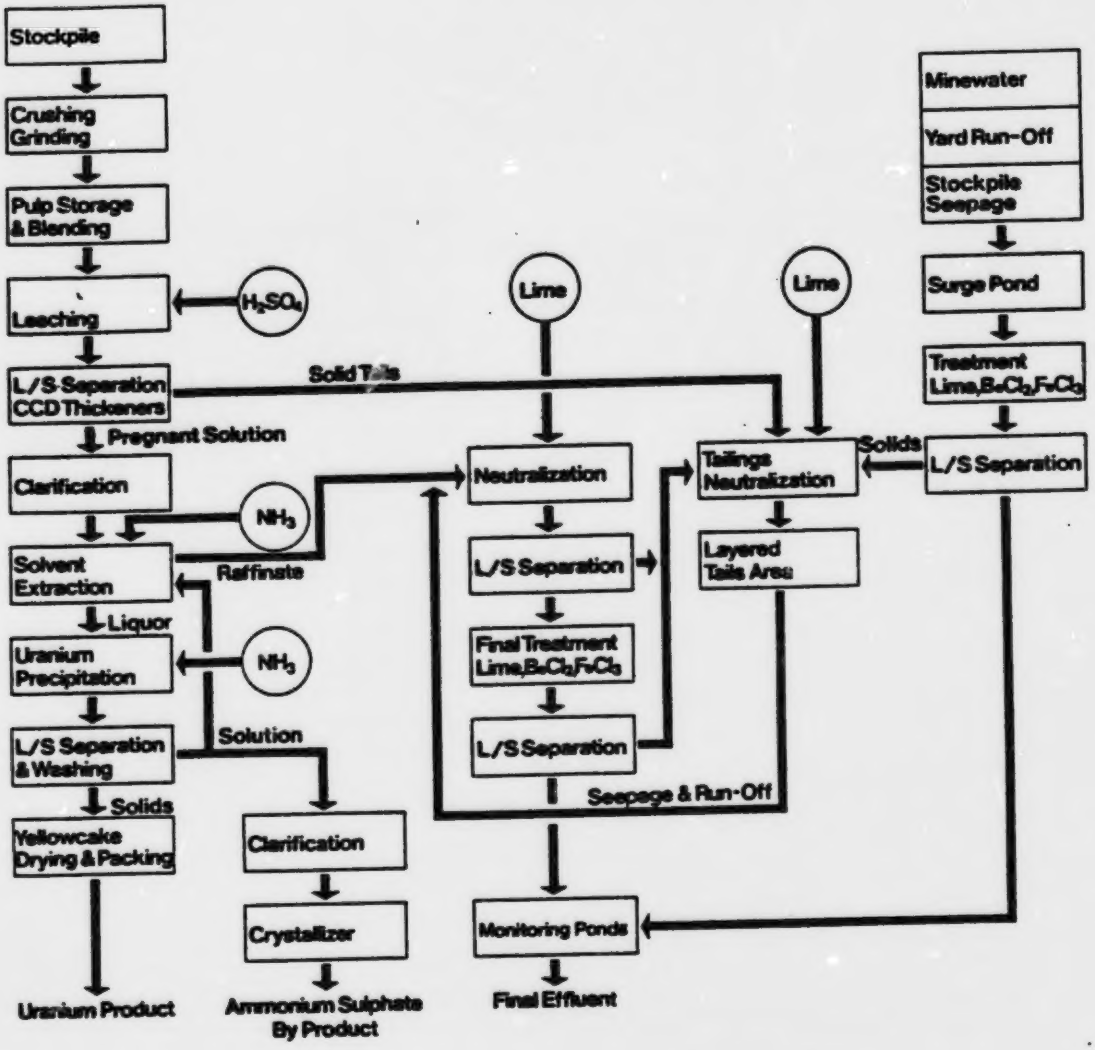

Stockpile
Crushing Grinding
Pulp Storage & Blending
Leaching
H₂SO₄
L/S Separation CCD Thickeners
Solid Tails
Pregnant Solution
Clarification
NH₃
Solvent Extraction
Raffinate
Liquor
Uranium Precipitation
NH₃
Solution
L/S Separation & Washing
Solids
Yellowcake Drying & Packing
Uranium Product
Clarification
Crystallizer
Ammonium Sulphate By Product
Lime
Neutralization
L/S Separation
Final Treatment Lime,BaCl₂,FeCl₃
L/S Separation
Seepage & Run-Off
Monitoring Ponds
Final Effluent
Lime
Tailings Neutralization
Layered Tails Area
Solids
Minewater
Yard Run-Off
Stockpile Seepage
Surge Pond
Treatment Lime,BaCl₂,FeCl₃
L/S Separation

Figure 3.1
Process Flowsheet
Midwest Project

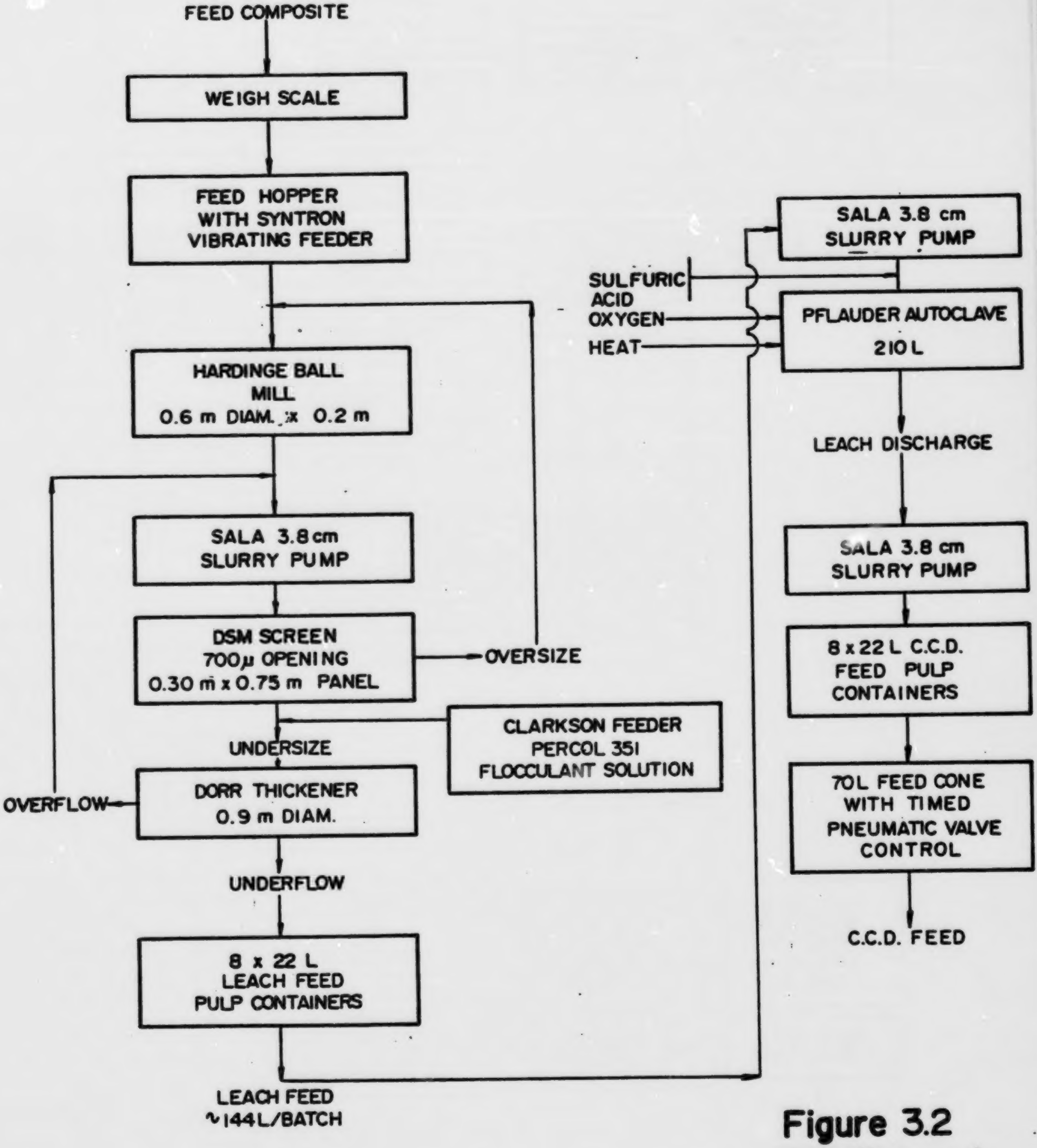

FEED COMPOSITE
WEIGH SCALE
FEED HOPPER WITH SYNTRON VIBRATING FEEDER
HARDINGE BALL MILL 0.6 m DIAM. x 0.2 m
SALA 3.8 cm SLURRY PUMP
DSM SCREEN 700μ OPENING 0.30 m x 0.75 m PANEL
OVERSIZE
CLARKSON FEEDER PERCOL 351 FLOCCULANT SOLUTION
UNDERSIZE
OVERFLOW
DORR THICKENER 0.9 m DIAM.
UNDERFLOW
8 x 22 L LEACH FEED PULP CONTAINERS
LEACH FEED ~144 L/BATCH
SALA 3.8 cm SLURRY PUMP
SULFURIC ACID
OXYGEN
HEAT
PFLAUDER AUTOCLAVE 210 L
LEACH DISCHARGE
SALA 3.8 cm SLURRY PUMP
8 x 22 L C.C.D. FEED PULP CONTAINERS
70 L FEED CONE WITH TIMED PNEUMATIC VALVE CONTROL
C.C.D. FEED

Figure 3.2 Grinding and Leaching

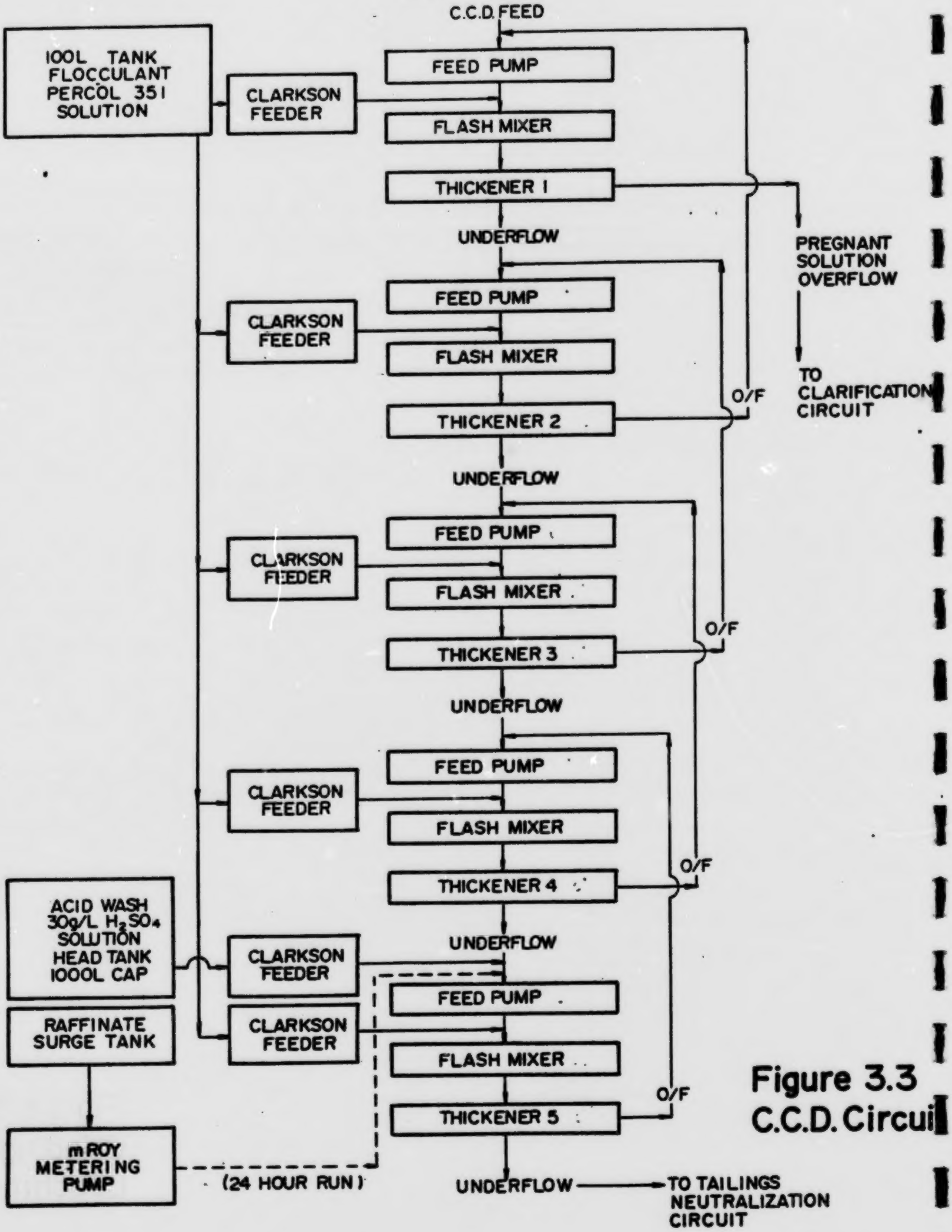

C.C.D. FEED
100L TANK FLOCCULANT PERCOL 351 SOLUTION
CLARKSON FEEDER
FEED PUMP
FLASH MIXER
THICKENER 1
UNDERFLOW
PREGNANT SOLUTION OVERFLOW
TO CLARIFICATION CIRCUIT
O/F
CLARKSON FEEDER
FEED PUMP
FLASH MIXER
THICKENER 2
UNDERFLOW
CLARKSON FEEDER
FEED PUMP
FLASH MIXER
THICKENER 3
O/F
UNDERFLOW
CLARKSON FEEDER
FEED PUMP
FLASH MIXER
THICKENER 4
O/F
UNDERFLOW
ACID WASH 30g/L H2SO4 SOLUTION HEAD TANK 1000L CAP
CLARKSON FEEDER
FEED PUMP
RAFFINATE SURGE TANK
CLARKSON FEEDER
FLASH MIXER
O/F
THICKENER 5
mROY METERING PUMP
(24 HOUR RUN)
UNDERFLOW
TO TAILINGS NEUTRALIZATION CIRCUIT

Figure 3.3
C.C.D. Circui

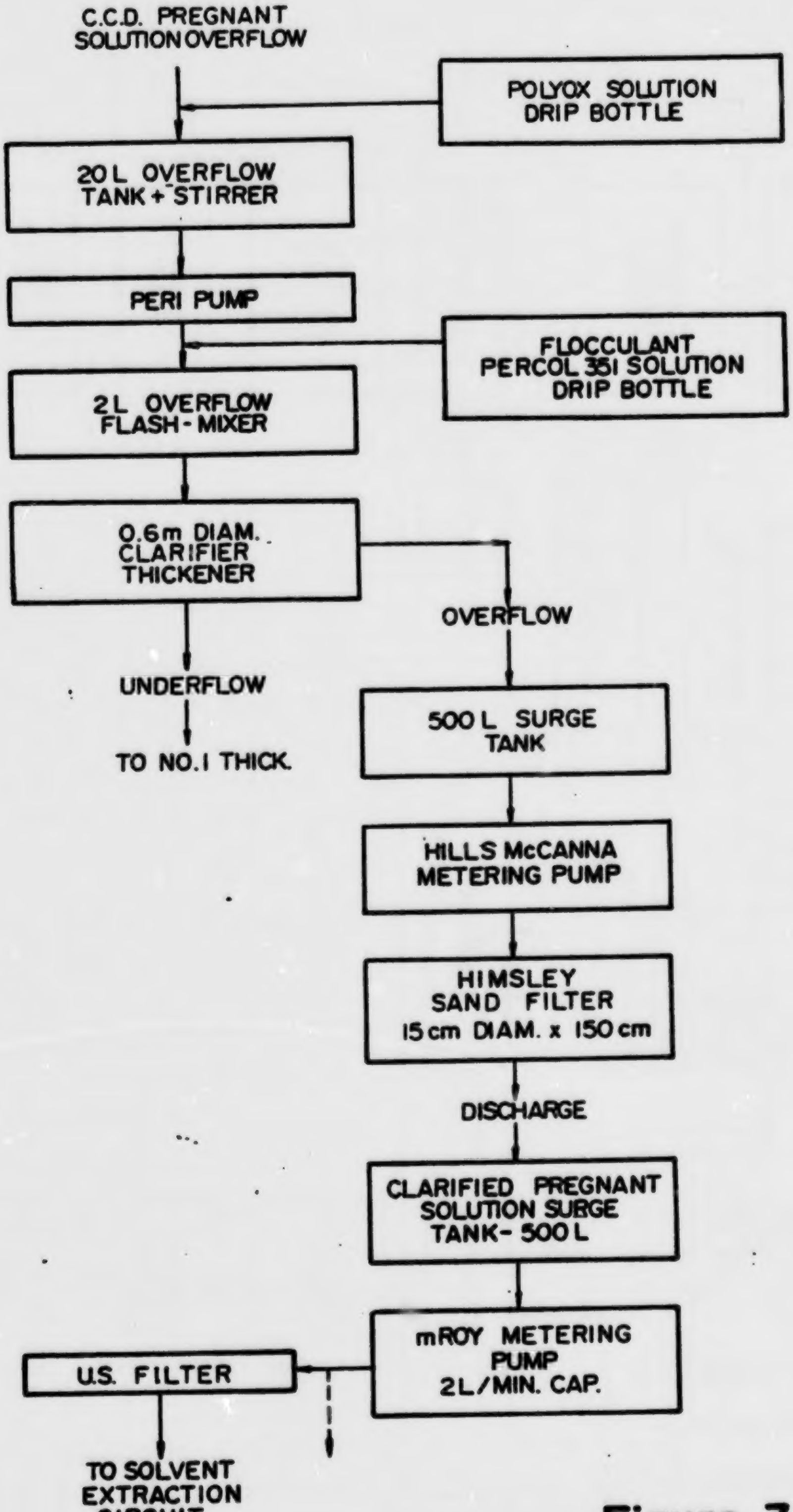

C.C.D. PREGNANT SOLUTION OVERFLOW
POLYOX SOLUTION DRIP BOTTLE
20 L OVERFLOW TANK + STIRRER
PERI PUMP
FLOCCULANT PERCOL 351 SOLUTION DRIP BOTTLE
2 L OVERFLOW FLASH - MIXER
0.6 m DIAM. CLARIFIER THICKENER
OVERFLOW
UNDERFLOW
TO NO. 1 THICK.
500 L SURGE TANK
HILLS McCANNA METERING PUMP
HIMSLEY SAND FILTER 15 cm DIAM. x 150 cm
DISCHARGE
CLARIFIED PREGNANT SOLUTION SURGE TANK - 500 L
mROY METERING PUMP 2 L/MIN. CAP.
U.S. FILTER
TO SOLVENT EXTRACTION CIRCUIT

Figure 3.4
Clarification Circuit

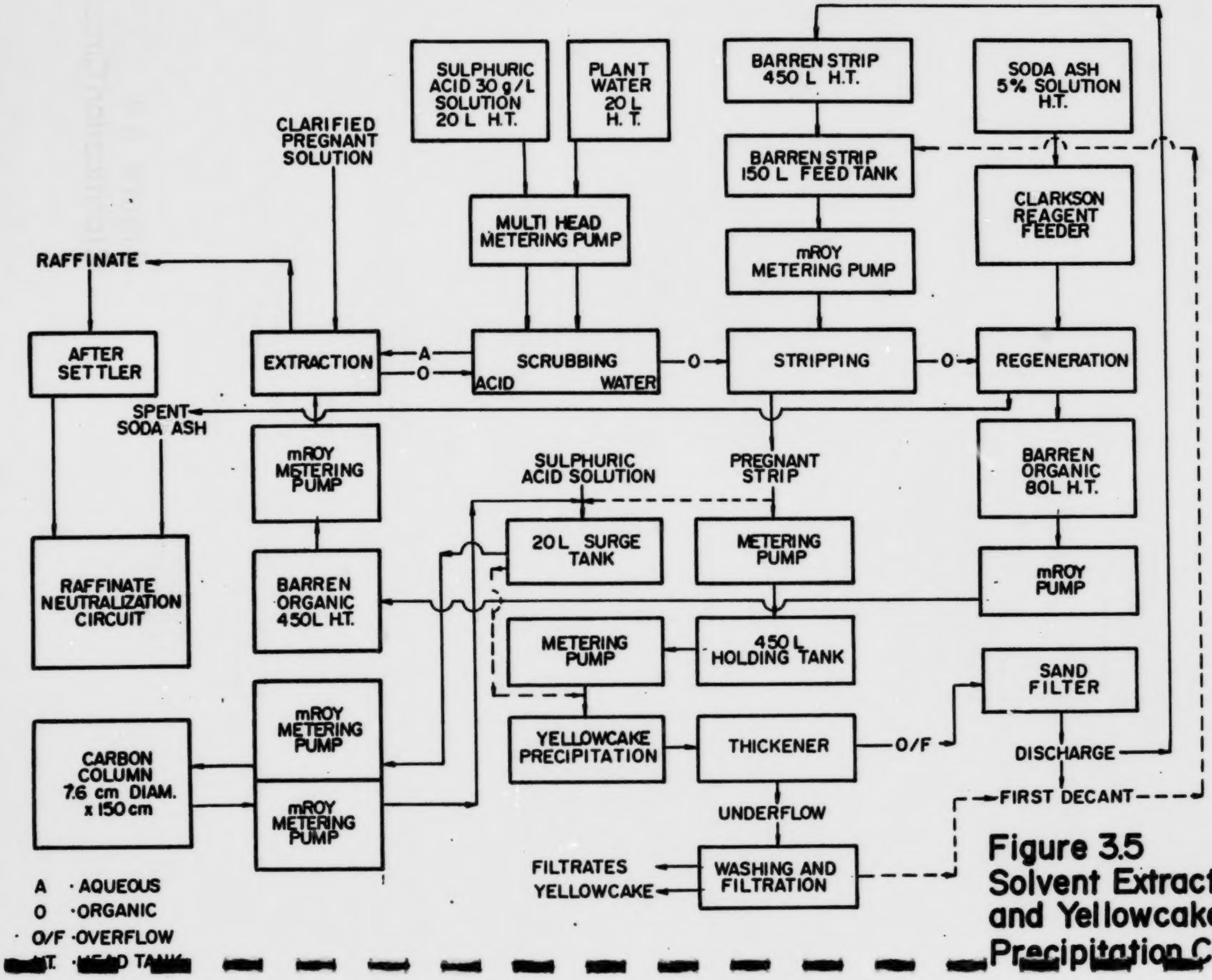

CLARIFIED PREGNANT SOLUTION
SULPHURIC ACID 30 g/L SOLUTION 20 L H.T.
PLANT WATER 20 L H. T.
MULTI HEAD METERING PUMP
BARREN STRIP 450 L H.T.
BARREN STRIP 150 L FEED TANK
mROY METERING PUMP
SODA ASH 5% SOLUTION H.T.
CLARKSON REAGENT FEEDER
RAFFINATE
AFTER SETTLER
EXTRACTION
A
O
SCRUBBING
ACID
WATER
O
STRIPPING
O
REGENERATION
SPENT SODA ASH
mROY METERING PUMP
SULPHURIC ACID SOLUTION
PREGNANT STRIP
BARREN ORGANIC 80L H.T.
20 L SURGE TANK
METERING PUMP
mROY PUMP
RAFFINATE NEUTRALIZATION CIRCUIT
BARREN ORGANIC 450L H.T.
METERING PUMP
450 L HOLDING TANK
SAND FILTER
mROY METERING PUMP
YELLOWCAKE PRECIPITATION
THICKENER
O/F
DISCHARGE
CARBON COLUMN 7.6 cm DIAM. x 150 cm
mROY METERING PUMP
FIRST DECANT
UNDERFLOW
FILTRATES
YELLOWCAKE
WASHING AND FILTRATION
A · AQUEOUS
O · ORGANIC
O/F · OVERFLOW
H.T. · HEAD TANK

Figure 3.5
Solvent Extraction
and Yellowcake
Precipitation Circuit

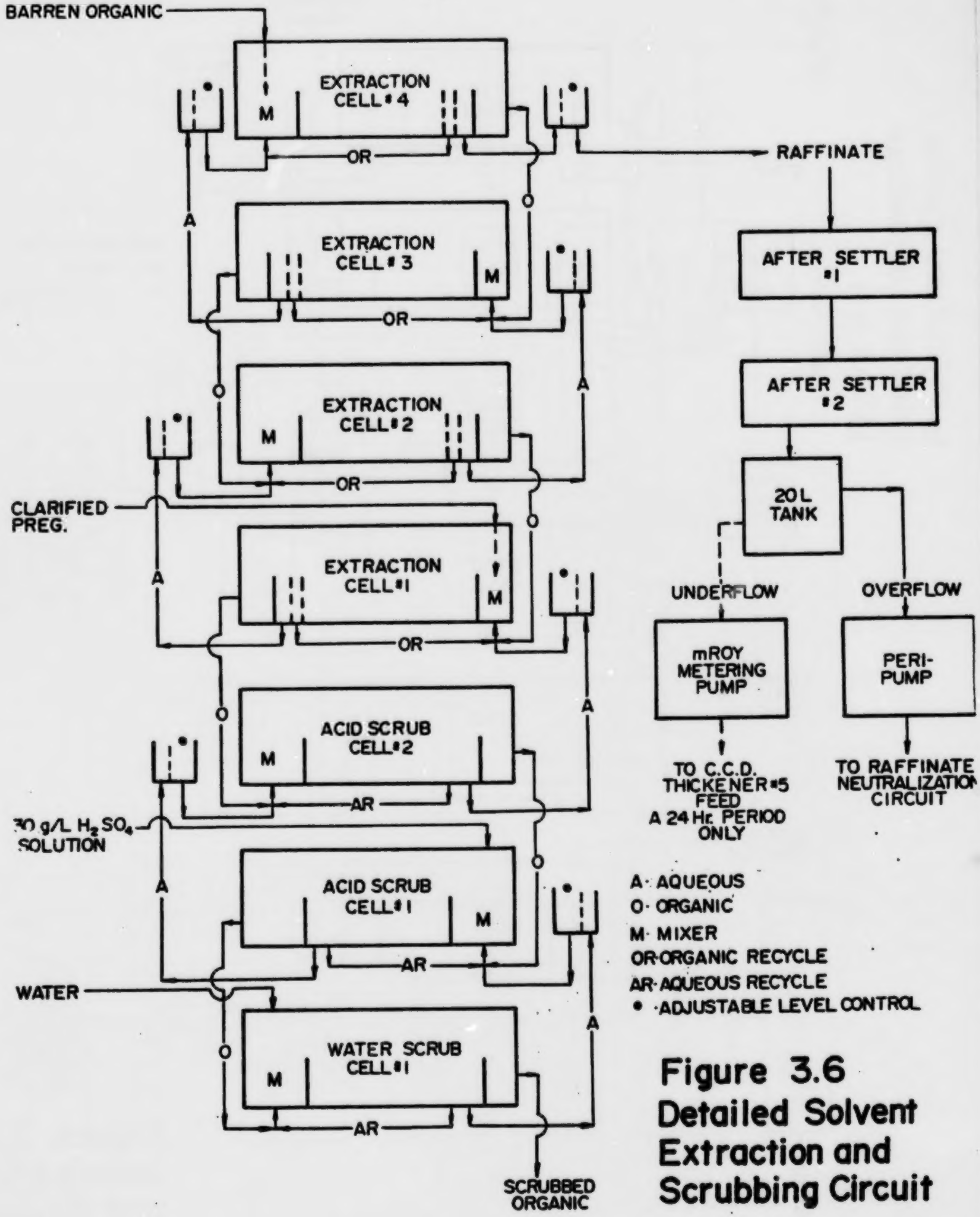

BARREN ORGANIC
EXTRACTION CELL #4
M
OR
RAFFINATE
A
O
EXTRACTION CELL #3
M
OR
AFTER SETTLER #1
AFTER SETTLER #2
EXTRACTION CELL #2
M
OR
A
20 L TANK
CLARIFIED PREG.
EXTRACTION CELL #1
M
OR
UNDERFLOW
OVERFLOW
mROY METERING PUMP
PERI-PUMP
ACID SCRUB CELL #2
M
AR
TO C.C.D. THICKENER #5 FEED A 24 Hr. PERIOD ONLY
TO RAFFINATE NEUTRALIZATION CIRCUIT
30 g/L H_2SO_4 SOLUTION
ACID SCRUB CELL #1
M
AR
A: AQUEOUS
O: ORGANIC
M: MIXER
OR: ORGANIC RECYCLE
AR: AQUEOUS RECYCLE
• : ADJUSTABLE LEVEL CONTROL
WATER
WATER SCRUB CELL #1
M
AR
SCRUBBED ORGANIC

Figure 3.6
Detailed Solvent Extraction and Scrubbing Circuit

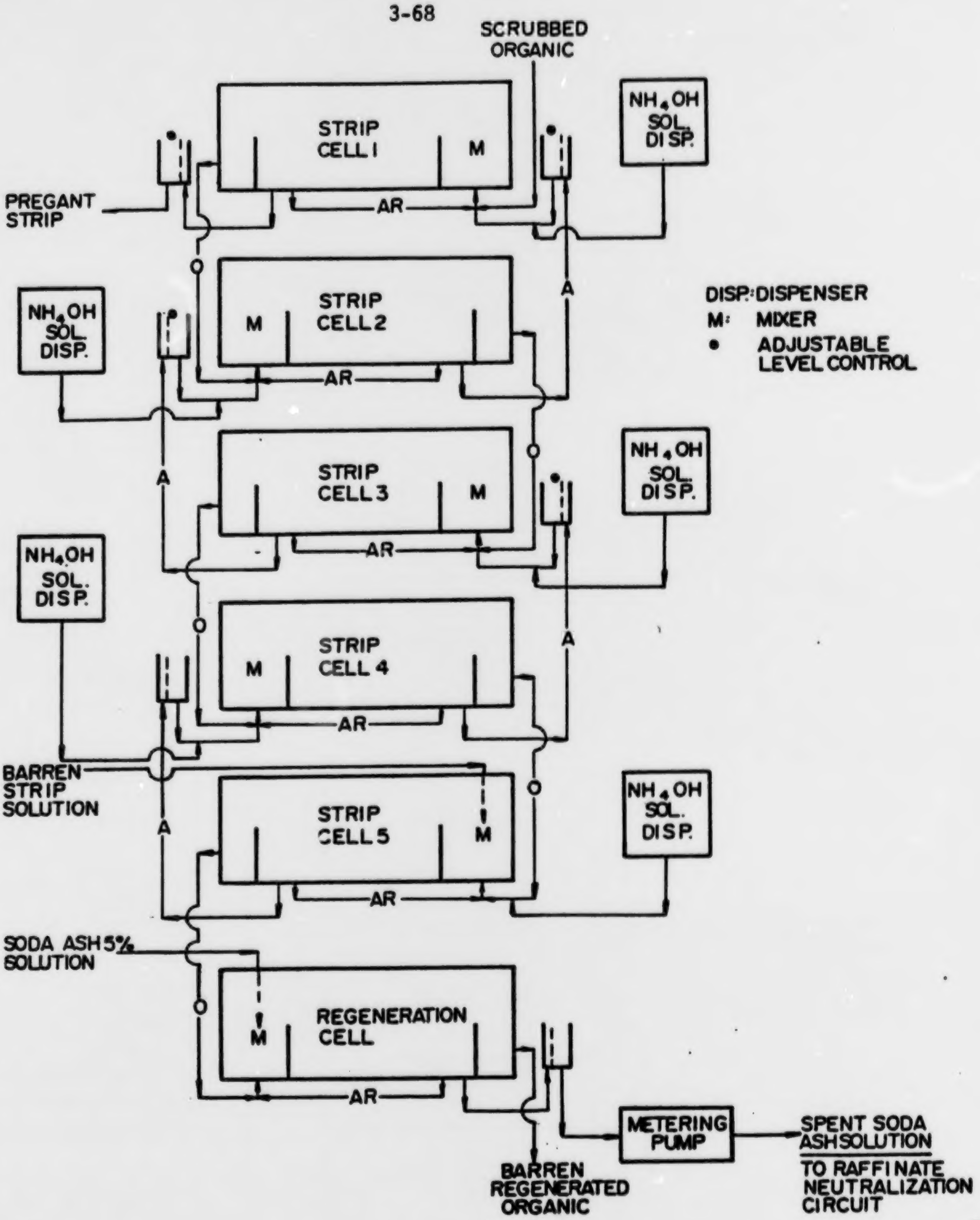

SCRUBBED ORGANIC
NH4OH SOL. DISP.
STRIP CELL 1
M
PREGANT STRIP
AR
O
A
DISP: DISPENSER
M: MIXER
• ADJUSTABLE LEVEL CONTROL
NH4OH SOL. DISP.
STRIP CELL 2
M
AR
STRIP CELL 3
M
NH4OH SOL. DISP.
A
AR
NH4OH SOL. DISP.
STRIP CELL 4
M
AR
BARREN STRIP SOLUTION
STRIP CELL 5
M
NH4OH SOL. DISP.
AR
SODA ASH 5% SOLUTION
REGENERATION CELL
M
AR
METERING PUMP
SPENT SODA ASH SOLUTION
TO RAFFINATE NEUTRALIZATION CIRCUIT
BARREN REGENERATED ORGANIC

Figure 3.7
Detailed Strippin
and Regeneratio
Circuit

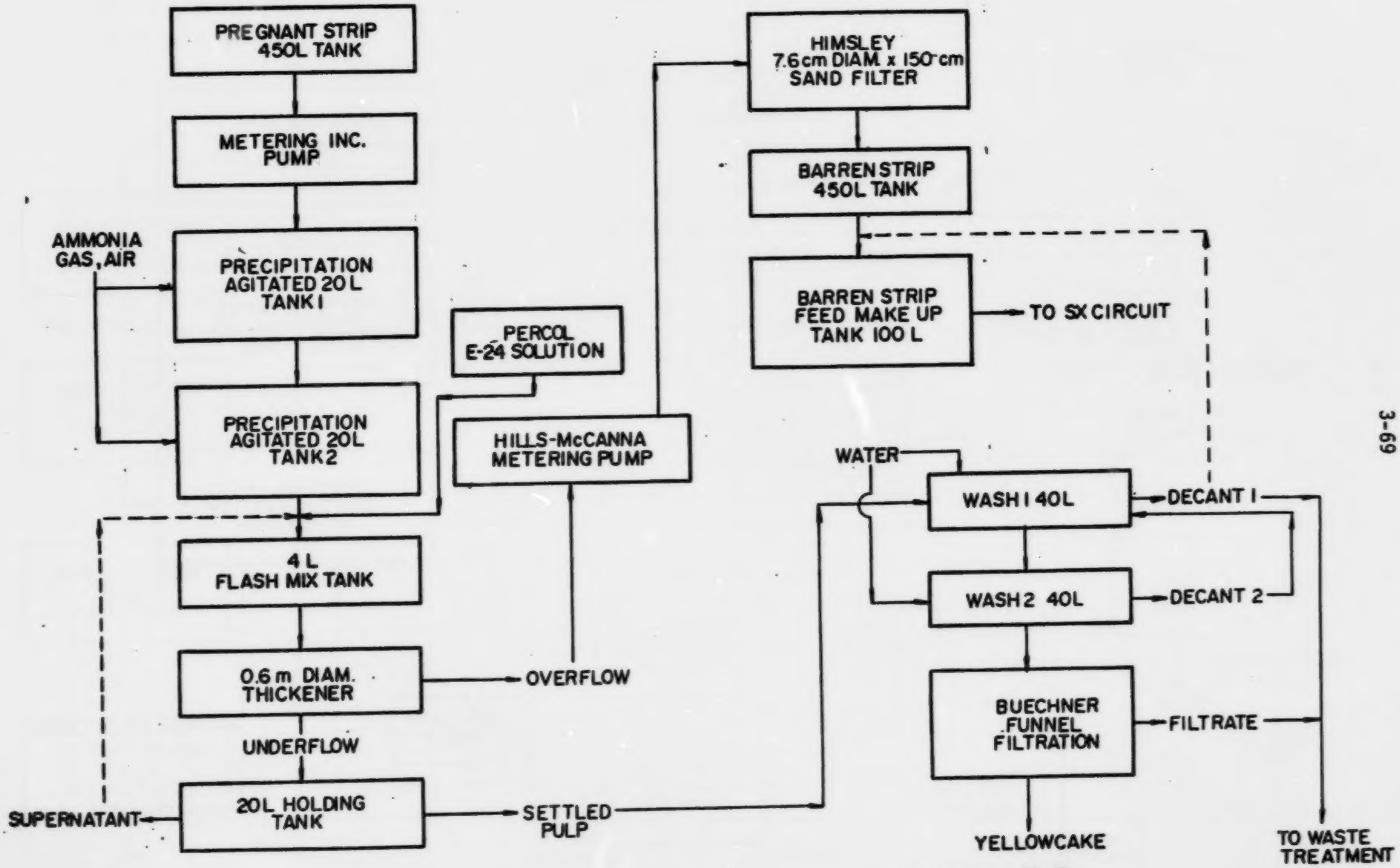

PREGNANT STRIP 450L TANK
METERING INC. PUMP
AMMONIA GAS, AIR
PRECIPITATION AGITATED 20 L TANK 1
PRECIPITATION AGITATED 20L TANK 2
PERCOL E-24 SOLUTION
4 L FLASH MIX TANK
0.6 m DIAM. THICKENER
UNDERFLOW
20L HOLDING TANK
SUPERNATANT
OVERFLOW
HILLS-McCANNA METERING PUMP
SETTLED PULP
HIMSLEY 7.6 cm DIAM x 150 cm SAND FILTER
BARREN STRIP 450L TANK
BARREN STRIP FEED MAKE UP TANK 100 L
TO SX CIRCUIT
WATER
WASH 1 40L
DECANT 1
WASH 2 40L
DECANT 2
BUECHNER FUNNEL FILTRATION
FILTRATE
YELLOWCAKE
TO WASTE TREATMENT

Figure 3.8
Detailed Yellowcake
Precipitation Circuit

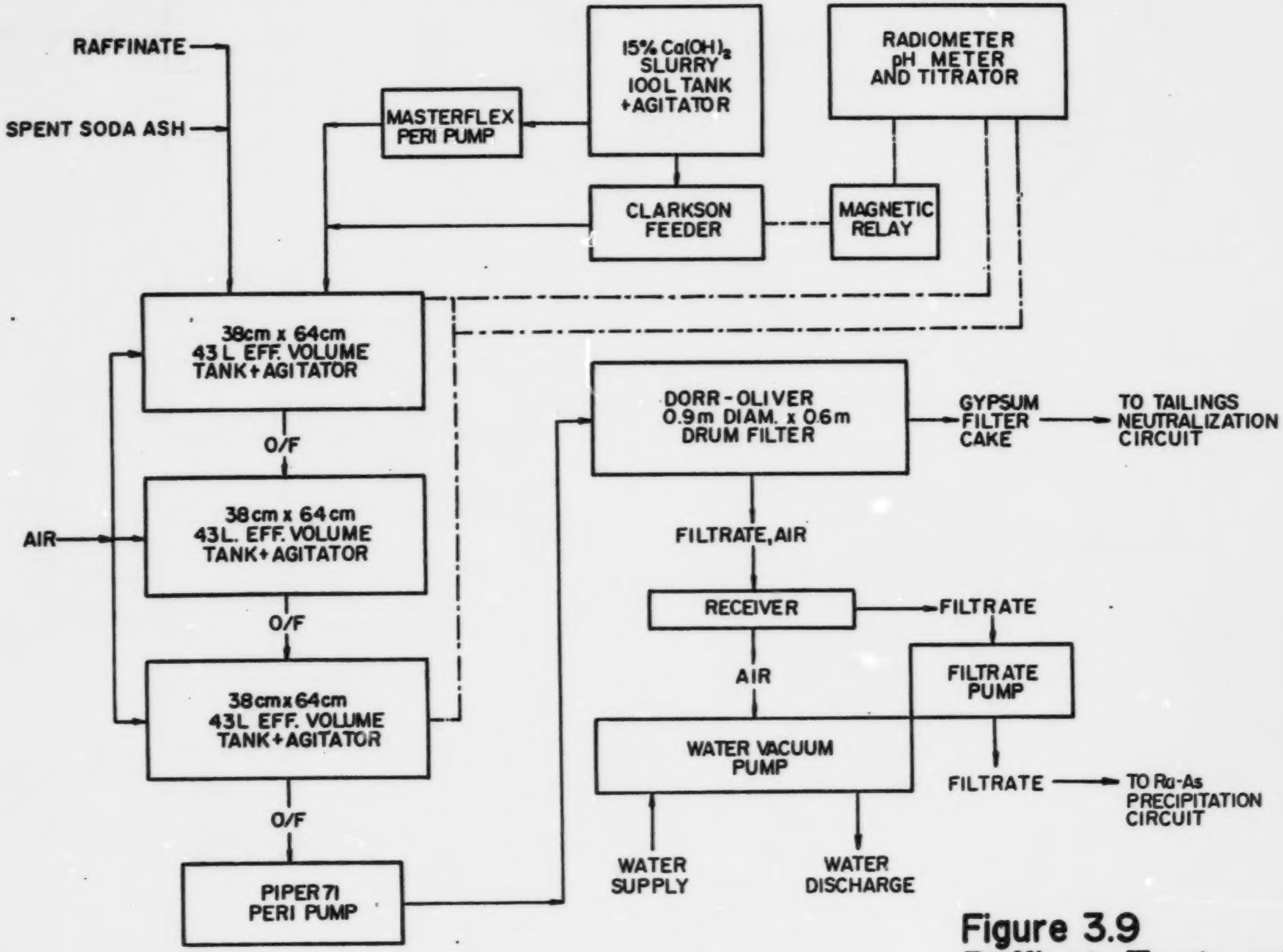

RAFFINATE
SPENT SODA ASH
MASTERFLEX PERI PUMP
15% Ca(OH)$_2$ SLURRY 100L TANK +AGITATOR
RADIOMETER pH METER AND TITRATOR
CLARKSON FEEDER
MAGNETIC RELAY
38cm x 64cm 43 L EFF. VOLUME TANK+AGITATOR
O/F
AIR
38cm x 64cm 43L. EFF. VOLUME TANK+AGITATOR
O/F
38cm x 64cm 43L EFF. VOLUME TANK+AGITATOR
O/F
PIPER 71 PERI PUMP
DORR-OLIVER 0.9m DIAM. x 0.6m DRUM FILTER
GYPSUM FILTER CAKE
TO TAILINGS NEUTRALIZATION CIRCUIT
FILTRATE, AIR
RECEIVER
FILTRATE
AIR
FILTRATE PUMP
WATER VACUUM PUMP
FILTRATE
TO Ra-As PRECIPITATION CIRCUIT
WATER SUPPLY
WATER DISCHARGE

Figure 3.9
Raffinate Treatment
Circuit

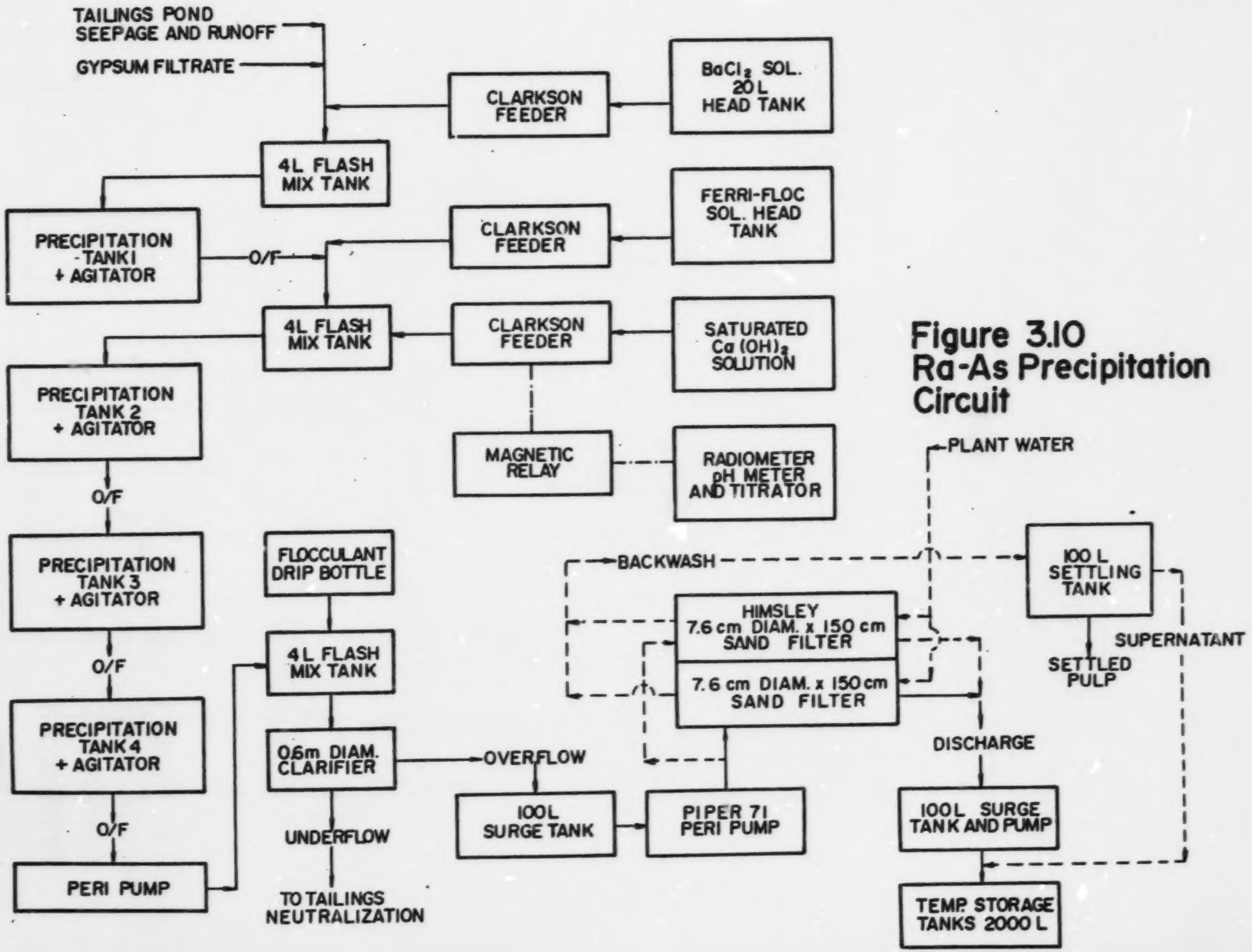

TAILINGS POND SEEPAGE AND RUNOFF
GYPSUM FILTRATE
BaCl₂ SOL. 20 L HEAD TANK
CLARKSON FEEDER
4L FLASH MIX TANK
PRECIPITATION TANK 1 + AGITATOR
O/F
FERRI-FLOC SOL. HEAD TANK
CLARKSON FEEDER
SATURATED Ca (OH)₂ SOLUTION
CLARKSON FEEDER
4L FLASH MIX TANK
PRECIPITATION TANK 2 + AGITATOR
MAGNETIC RELAY
RADIOMETER pH METER AND TITRATOR
O/F
PRECIPITATION TANK 3 + AGITATOR
O/F
PRECIPITATION TANK 4 + AGITATOR
O/F
PERI PUMP
FLOCCULANT DRIP BOTTLE
4L FLASH MIX TANK
0.6m DIAM. CLARIFIER
UNDERFLOW
TO TAILINGS NEUTRALIZATION
OVERFLOW
100L SURGE TANK
PIPER 71 PERI PUMP
BACKWASH
HIMSLEY 7.6 cm DIAM. x 150 cm SAND FILTER
7.6 cm DIAM. x 150 cm SAND FILTER
PLANT WATER
100 L SETTLING TANK
SETTLED PULP
SUPERNATANT
DISCHARGE
100L SURGE TANK AND PUMP
TEMP. STORAGE TANKS 2000 L

Figure 3.10
Ra-As Precipitation Circuit

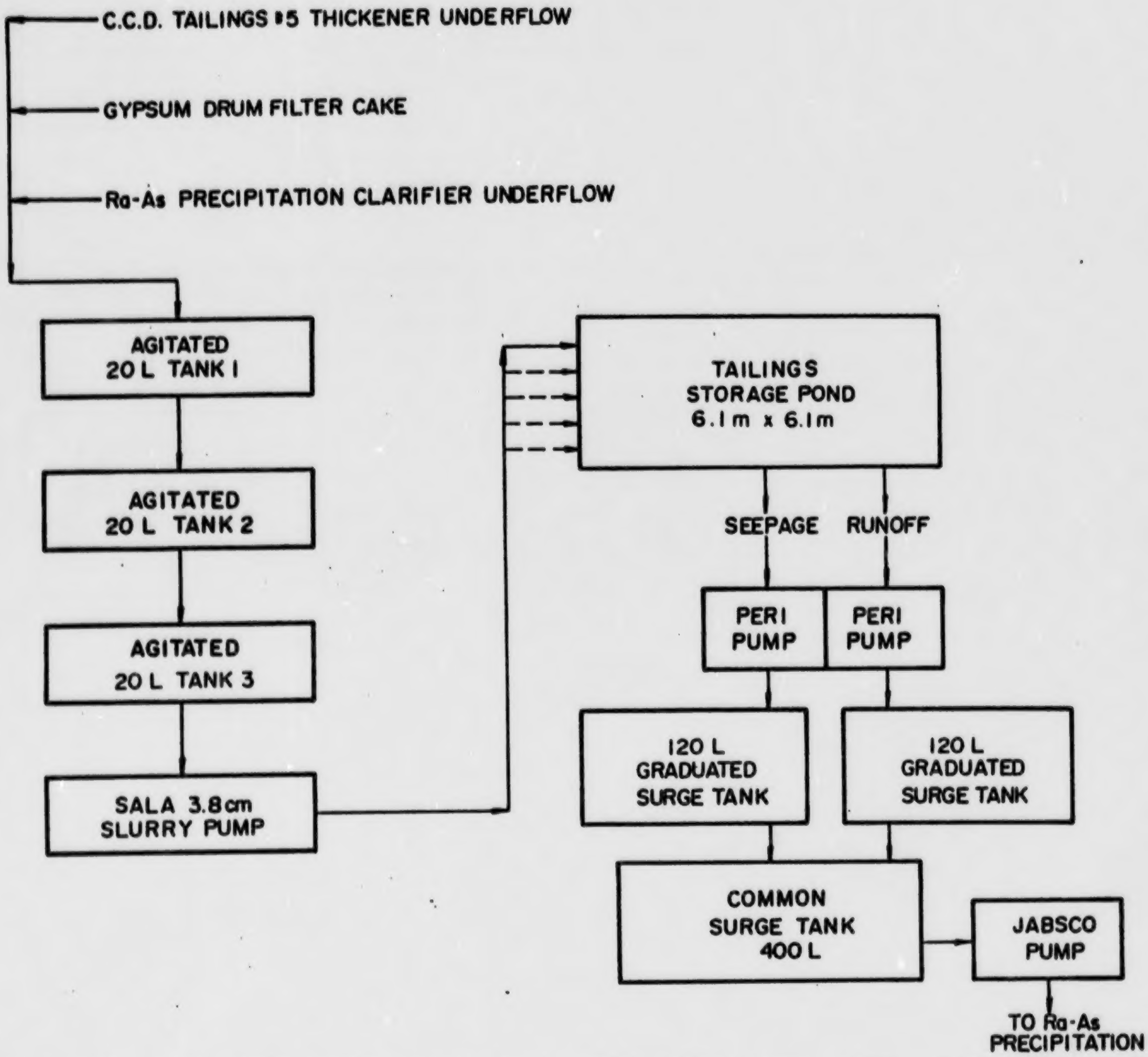

C.C.D. TAILINGS #5 THICKENER UNDERFLOW
GYPSUM DRUM FILTER CAKE
Ra-As PRECIPITATION CLARIFIER UNDERFLOW
AGITATED 20 L TANK 1
AGITATED 20 L TANK 2
AGITATED 20 L TANK 3
SALA 3.8 cm SLURRY PUMP
TAILINGS STORAGE POND 6.1 m x 6.1 m
SEEPAGE
RUNOFF
PERI PUMP
PERI PUMP
120 L GRADUATED SURGE TANK
120 L GRADUATED SURGE TANK
COMMON SURGE TANK 400 L
JABSCO PUMP
TO Ra-As PRECIPITATION

Figure 3.11
Tailings Treatment Circuit

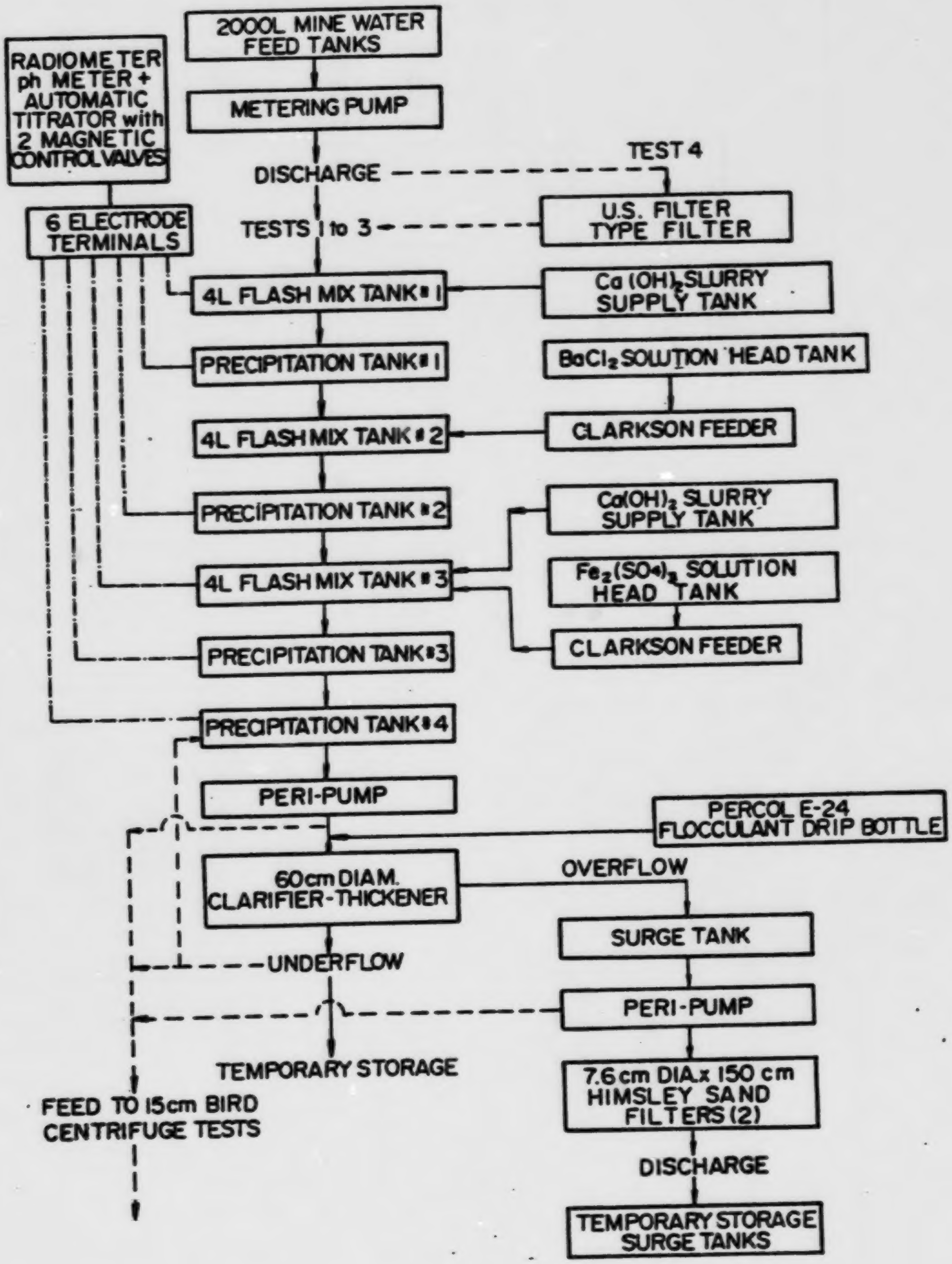
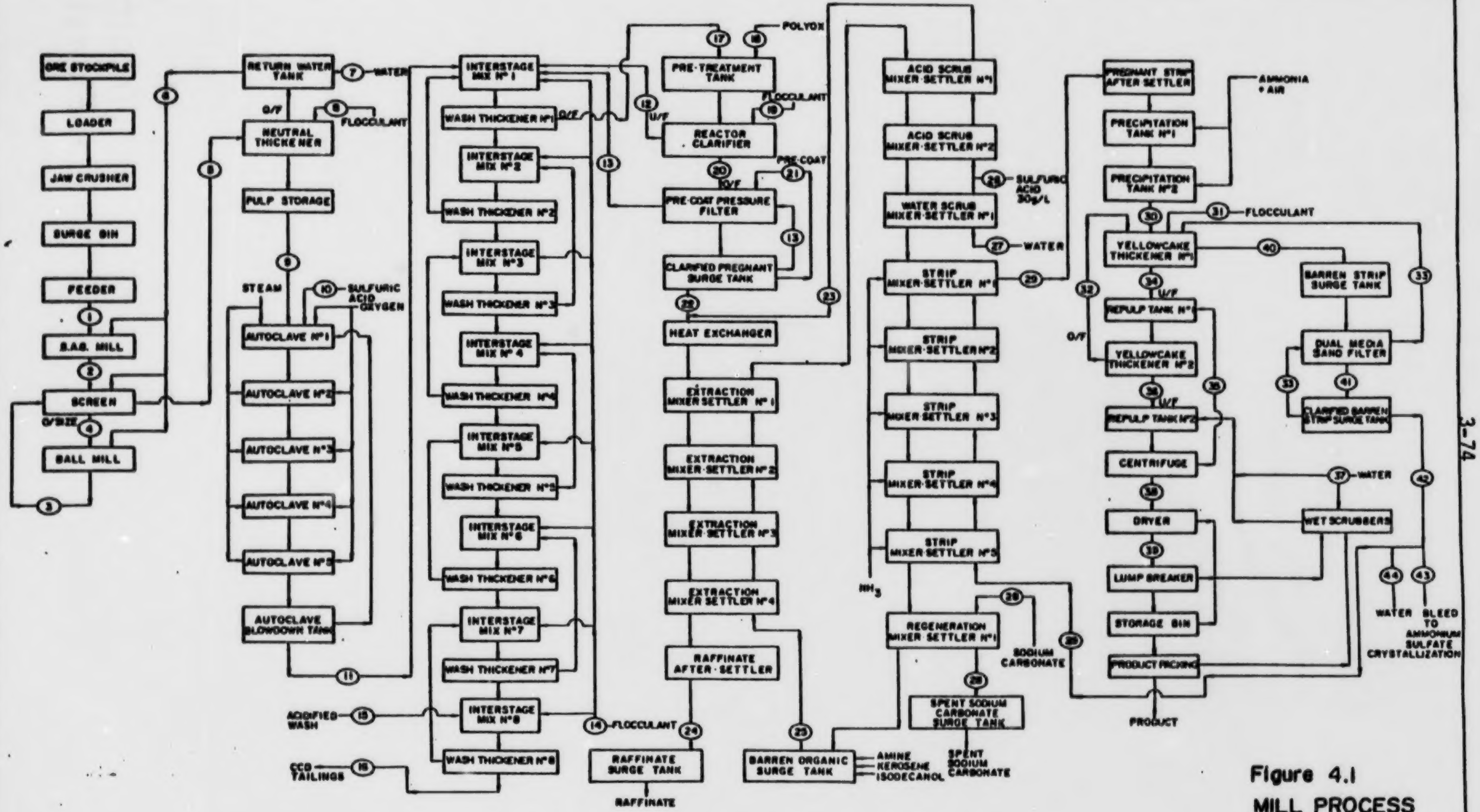

2000L MINE WATER FEED TANKS
METERING PUMP
RADIOMETER ph METER + AUTOMATIC TITRATOR with 2 MAGNETIC CONTROL VALVES
6 ELECTRODE TERMINALS
DISCHARGE
TEST 4
TESTS 1 to 3
U.S. FILTER TYPE FILTER
4L FLASH MIX TANK # 1
Ca (OH)$_2$ SLURRY SUPPLY TANK
PRECIPITATION TANK # 1
BaCl$_2$ SOLUTION HEAD TANK
4L FLASH MIX TANK # 2
CLARKSON FEEDER
PRECIPITATION TANK # 2
Ca(OH)$_2$ SLURRY SUPPLY TANK
4L FLASH MIX TANK # 3
Fe$_2$(SO$_4$)$_3$ SOLUTION HEAD TANK
PRECIPITATION TANK # 3
CLARKSON FEEDER
PRECIPITATION TANK # 4
PERI-PUMP
PERCOL E-24 FLOCCULANT DRIP BOTTLE
60cm DIAM. CLARIFIER-THICKENER
OVERFLOW
SURGE TANK
UNDERFLOW
PERI-PUMP
TEMPORARY STORAGE
FEED TO 15cm BIRD CENTRIFUGE TESTS
7.6 cm DIA.x 150 cm HIMSLEY SAND FILTERS (2)
DISCHARGE
TEMPORARY STORAGE SURGE TANKS

Figure 3.12
Mine Water
Treatment
Flowsheet


ORE STOCKPILE
LOADER
JAW CRUSHER
SURGE BIN
FEEDER
S.A.G. MILL
SCREEN
O/SIZE
BALL MILL
RETURN WATER TANK
WATER
O/F
NEUTRAL THICKENER
FLOCCULANT
PULP STORAGE
STEAM
SULFURIC ACID
OXYGEN
AUTOCLAVE Nº1
AUTOCLAVE Nº2
AUTOCLAVE Nº3
AUTOCLAVE Nº4
AUTOCLAVE Nº5
AUTOCLAVE BLOWDOWN TANK
ACIDIFIED WASH
CCD TAILINGS
INTERSTAGE MIX Nº1
WASH THICKENER Nº1
INTERSTAGE MIX Nº2
WASH THICKENER Nº2
INTERSTAGE MIX Nº3
WASH THICKENER Nº3
INTERSTAGE MIX Nº4
WASH THICKENER Nº4
INTERSTAGE MIX Nº5
WASH THICKENER Nº5
INTERSTAGE MIX Nº6
WASH THICKENER Nº6
INTERSTAGE MIX Nº7
WASH THICKENER Nº7
INTERSTAGE MIX Nº8
WASH THICKENER Nº8
O/F
U/F
FLOCCULANT
POLYOX
PRE-TREATMENT TANK
FLOCCULANT
REACTOR CLARIFIER
O/F
PRE-COAT
PRE-COAT PRESSURE FILTER
CLARIFIED PREGNANT SURGE TANK
HEAT EXCHANGER
EXTRACTION MIXER-SETTLER Nº1
EXTRACTION MIXER-SETTLER Nº2
EXTRACTION MIXER-SETTLER Nº3
EXTRACTION MIXER-SETTLER Nº4
RAFFINATE AFTER-SETTLER
RAFFINATE SURGE TANK
RAFFINATE
BARREN ORGANIC SURGE TANK
AMINE
KEROSENE
ISODECANOL
ACID SCRUB MIXER-SETTLER Nº1
ACID SCRUB MIXER-SETTLER Nº2
SULFURIC ACID 30g/L
WATER SCRUB MIXER-SETTLER Nº1
WATER
STRIP MIXER-SETTLER Nº1
STRIP MIXER-SETTLER Nº2
STRIP MIXER-SETTLER Nº3
STRIP MIXER-SETTLER Nº4
STRIP MIXER-SETTLER Nº5
NH3
REGENERATION MIXER-SETTLER Nº1
SODIUM CARBONATE
SPENT SODIUM CARBONATE SURGE TANK
SPENT SODIUM CARBONATE
PREGNANT STRIP AFTER SETTLER
AMMONIA + AIR
PRECIPITATION TANK Nº1
PRECIPITATION TANK Nº2
FLOCCULANT
YELLOWCAKE THICKENER Nº1
U/F
REPULP TANK Nº1
O/F
YELLOWCAKE THICKENER Nº2
U/F
REPULP TANK Nº2
CENTRIFUGE
DRYER
LUMP BREAKER
STORAGE BIN
PRODUCT PACKING
PRODUCT
BARREN STRIP SURGE TANK
DUAL MEDIA SAND FILTER
CLARIFIED BARREN STRIP SURGE TANK
WATER
WET SCRUBBERS
WATER
BLEED TO AMMONIUM SULFATE CRYSTALLIZATION
1
2
3
4
5
6
7
8
9
10
11
12
13
13
14
15
16
17
18
19
20
21
22
23
23
24
25
26
27
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29
30
31
32
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43
44

Figure 4.1
MILL PROCESS
FLOWSHEET

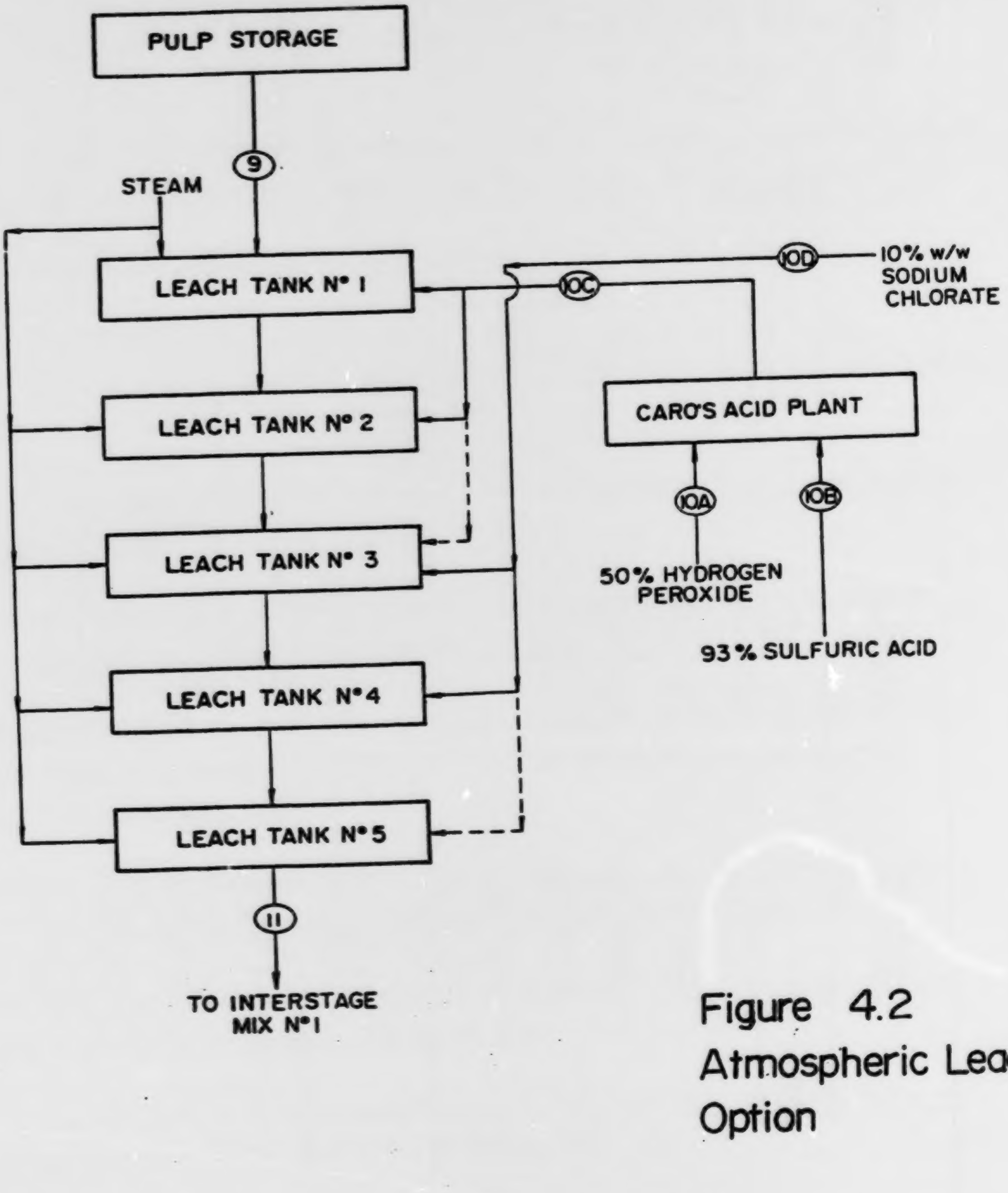
PULP STORAGE
9
STEAM
LEACH TANK N° 1
LEACH TANK N° 2
LEACH TANK N° 3
LEACH TANK N° 4
LEACH TANK N° 5
11
TO INTERSTAGE MIX N° 1
10D
10% w/w SODIUM CHLORATE
10C
CARO'S ACID PLANT
10A
10B
50% HYDROGEN PEROXIDE
93% SULFURIC ACID

Figure 4.2
Atmospheric Leaching
Option

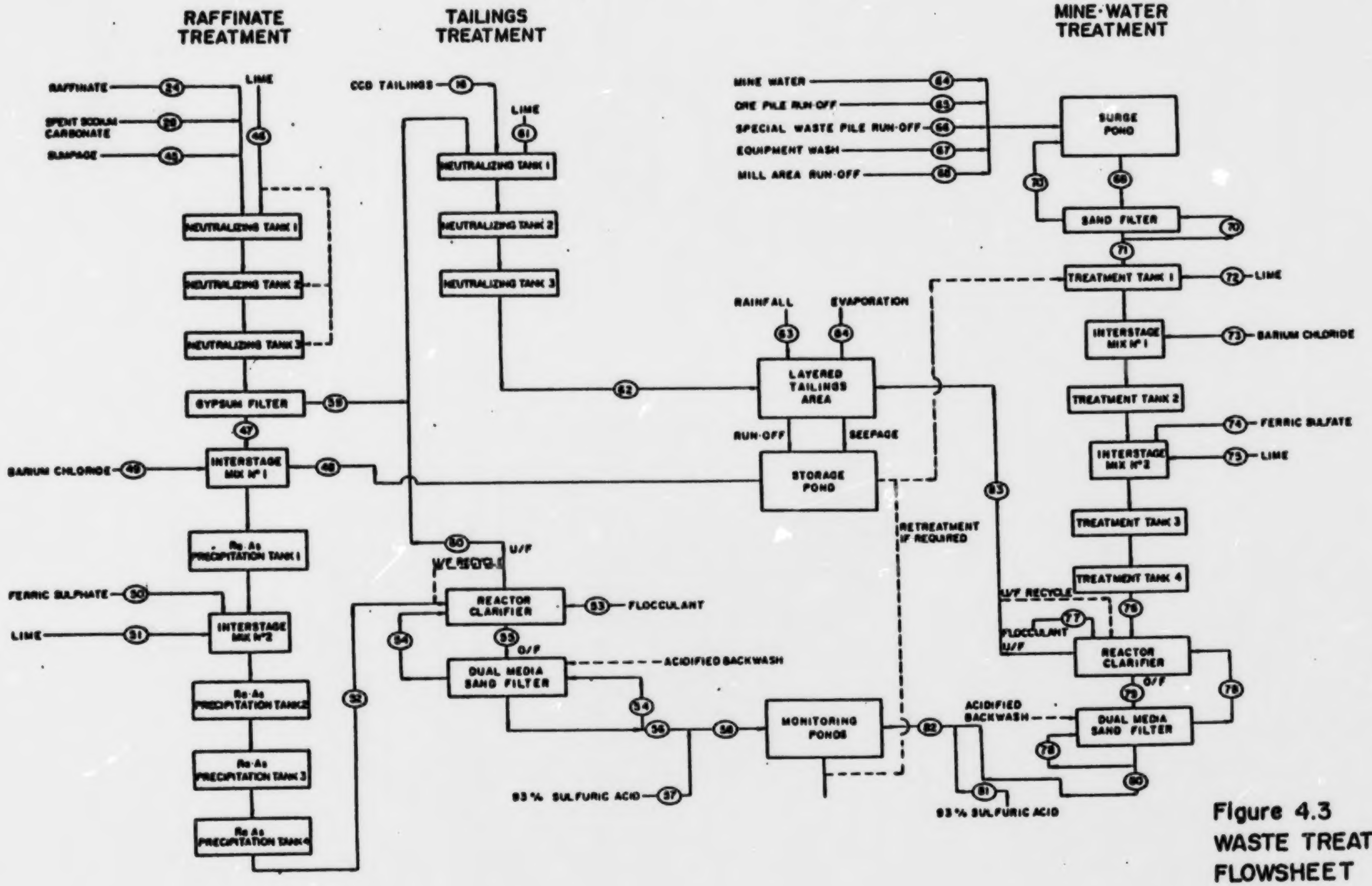
RAFFINATE TREATMENT
TAILINGS TREATMENT
MINE·WATER TREATMENT
RAFFINATE
24
SPENT SODIUM CARBONATE
29
SEEPAGE
45
LIME
46
NEUTRALIZING TANK 1
NEUTRALIZING TANK 2
NEUTRALIZING TANK 3
GYPSUM FILTER
39
47
BARIUM CHLORIDE
49
INTERSTAGE MIX Nº 1
48
Re·As PRECIPITATION TANK 1
FERRIC SULPHATE
50
LIME
51
INTERSTAGE MIX Nº2
Re·As PRECIPITATION TANK 2
Re·As PRECIPITATION TANK 3
Re·As PRECIPITATION TANK 4
52
CCD TAILINGS
16
LIME
61
NEUTRALIZING TANK 1
NEUTRALIZING TANK 2
NEUTRALIZING TANK 3
62
50
U/F
U/F RECYCLE
REACTOR CLARIFIER
53
FLOCCULANT
55
O/F
54
DUAL MEDIA SAND FILTER
ACIDIFIED BACKWASH
54
56
58
93 % SULFURIC ACID
57
MONITORING PONDS
RAINFALL
EVAPORATION
63
64
LAYERED TAILINGS AREA
RUN·OFF
SEEPAGE
STORAGE POND
RETREATMENT IF REQUIRED
MINE WATER
64
ORE PILE RUN·OFF
65
SPECIAL WASTE PILE RUN·OFF
66
EQUIPMENT WASH
67
MILL AREA RUN·OFF
68
SURGE POND
69
70
SAND FILTER
70
71
TREATMENT TANK 1
72
LIME
INTERSTAGE MIX Nº 1
73
BARIUM CHLORIDE
TREATMENT TANK 2
74
FERRIC SULFATE
INTERSTAGE MIX Nº2
75
LIME
TREATMENT TANK 3
TREATMENT TANK 4
83
U/F RECYCLE
76
77
FLOCCULANT
U/F
REACTOR CLARIFIER
O/F
79
78
ACIDIFIED BACKWASH
DUAL MEDIA SAND FILTER
79
80
82
81
93 % SULFURIC ACID

Figure 4.3
WASTE TREATMENT
FLOWSHEET

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DISCUSSION OF PAPER BY L. MELIS

P. Fournier	What is the screen size of the ore? Is there any difference between pressure and atmospheric leaching?
L. Melis	There was no change in conditions, the grind being 30% minus 200 mesh. The extraction efficiency was 98.5-99% by either route, using 8 stages of CCD. The expected soluble loss is 0.3%.
W. Milde	What type of pre-coat is used, and were there any problems encountered in solvent extraction?
L. Melis	The precoat used was Dicalite 436. There were some problems in solvent extraction. The temperature should be maintained at 35°C.
R. Rollwagen	Was any coagulant used in sand filtration?
L. Melis	No, but Polyox (2-5 ppm) was used, followed by Percol 351.
J. Goode	Was $NaClO_3$ considered?
L. Melis	Yes, the same extraction was achieved but a much longer time was required as compared to Caro's acid.

PRESENTATION TO CANADIAN URANIUM PRODUCERS MEETING AT BANCROFT
MAY, 1982

STATUS REPORT ON KIDD CREEK AQUITAINE PROJECT IN NOVA SCOTIA

R. John
Kidd Creek Mines

PRESENTATION TO CANADIAN URANIUM PRODUCERS MEETING AT BANCROFT

MAY, 1982

GORDON RITCEY HAS ASKED ME TO MAKE A PRESENTATION TODAY THAT WILL EXPLAIN WHAT IS HAPPENING IN NOVA SCOTIA. I WOULD LIKE TO DO THIS BY SPLITTING MY PRESENTATION INTO TWO PARTS. THE FIRST PART WILL BE THE HISTORY AND, IF THE GEOLOGISTS PROMISE NOT TO BE TOO CRITICAL, SOMETHING OF THE GEOLOGY OF THE AREA. THEN, I WOULD LIKE TO TALK ABOUT THE CURRENT SITUATION WITH THE MORATORIUM AND THE PUBLIC INQUIRY. SO FIRST LET US LOOK AT THE HISTORY AND THE GEOLOGY.

AS I HAVE SAID, I AM NOT A GEOLOGIST AND I WOULD ASK THOSE WHO ARE TECHNICALLY MORE COMPETENT IN THIS TO BARE WITH MY SIMPLIFIED INFORMATION.

SOME YEARS AGO, IT WAS PROPOSED THAT NOVA SCOTIA, INDEED, THE WHOLE MARITIME REGION WAS IN FACT PART OF EUROPE - NORTH AFRICAN GEOLOGY. A WIDELY ACCEPTED THEORY AMONGST GEOLOGISTS IS THAT THE MARITIME REGION OF CANADA WAS ONCE CO-EXISTANT WITH WESTERN EUROPE FORMING A SUPER-CONTINENT. IN THE GEOLOGICAL PAST OUR "CANADIAN" PORTION HAS DRIFTED AWAY FROM EUROPE. IF THIS IS THE CASE, THEN, THE GEOLOGY OF THIS AREA, WHICH ENCOMPASSES NEW ENGLAND, NEW BRUNSWICK, NOVA SCOTIA AND NEWFOUNDLAND, SHOULD HAVE VERY SIMILAR GEOLOGY TO FRANCE, SPAIN, AND PORTUGAL. WITH THIS SCIENTIFIC MODEL IN MIND, NOVA SCOTIA WAS RE-EXPLORED FROM A DIFFERENT VIEWPOINT. OUR GROUP BEING ORIGINALLY PART OF A FRENCH-OWNED COMPANY INCLUDED SOME FRENCH GEOLOGISTS WHO BROUGHT THEIR EUROPEAN EXPERIENCE WITH THEM. BECAUSE OF THESE CHANGES IN VIEWPOINT, THE GEOLOGISTS LOOKED AT NOVA SCOTIA IN A VERY DIFFERENT

WAY AND SEARCHED FOR DIFFERENT METALS IN DIFFERENT PLACES. IT WOULD APPEAR THAT THESE CONCEPTS ARE TRUE BECAUSE WE ARE NOW FINDING MINERAL DEPOSITS THAT ARE INDEED SIMILAR TO THE EUROPEAN ONES. OUR OWN GROUP HAS FOUND A MINERALIZED ZONE CONTAINING URANIUM, SHELL HAS A TIN DEPOSIT NEAR YARMOUTH, AND THERE HAVE BEEN A NUMBER OF SIMILAR INTERESTING SHOWINGS.

FOR US, THE PROCESS BEGAN WITH AN AERIAL RADIOMETRIC SURVEY IN 1976. THIS SHOWED AN ANOMALY AT THE AREA WE NOW CALL MILLET BROOK. THE SUBSEQUENT GROUND PROSPECTING IN 1977 UNEARTHED SOME METATORBERNITE IN THE TILL COVER. SIMPLE HAND DUG TRENCHES REVEALED OTHER URANIUM PHOSPHATE MINERALIZED ZONES. DURING THE SUMMER OF 1978, WE INCREASED OUR ACTIVITIES SIGNIFICANTLY. WE BEGAN A MAJOR TRENCHING PROGRAM. THE RESPONSE TO THIS WAS SUFFICIENTLY ENCOURAGING FOR US TO STAKE OVER 1700 NEW CLAIMS IN THE AREA. THE INITIAL DRILLING PROGRAM WAS STARTED IN THE WINTER OF 1978-1979 AND CONTINUED AT AN EVER INCREASING PACE UNTIL SEPTEMBER, 1981.

THE MILLET BROOK PROPERTY IS ABOUT 50 KM NORTHWEST OF HALIFAX, NOVA SCOTIA. IT LIES NEAR TWO SMALL COMMUNITIES KNOWN AS VAUGHANS AND LEMINSTER. THIS AREA HAS BEEN KNOWN AS A MINERALIZED DISTRICT FOR ALMOST A CENTURY. THERE ARE TIN, MOLYBDENUM, TUNGSTEN, AND MANGANESE DEPOSITS, SOME OF WHICH HAVE BEEN WORKED. ALL OF THESE DEPOSITS INCLUDING OUR OWN MILLET BROOK URANIUM MINERALIZATION OCCUR AS VEIN-TYPE DEPOSITS WITHIN THE GRANITES. THE MINERAL ZONES ARE CLOSE TO VERTICAL, WITH GRADES RUNNING BETWEEN 0.1 AND 0.2%. INTERESTINGLY THE LARGEST DEPOSIT WE HAVE SO FAR RESPONDED VERY POORLY TO RADIOMETRICS AND SOIL GEOCHEMISTRY. THIS IS BECAUSE OF THE 2-5 METERS OF GLACIAL TILL AND THE OVERLYING BOULDER

FIELD. FROM WHAT I HAVE SEEN OF THE ROCK SO FAR, APPEARS TO BE FAIRLY SOFT, EASILY FRACTURED, AND SHOULD BE RELATIVELY EASY TO MILL. WE ARE ALSO FORTUNATE WITH THE LACK OF SULFUR; TYPICALLY WE ESTIMATE LESS THAN ONE PERCENT PYRITE.

AT THIS STAGE, OF COURSE, WE HAVE NOT DEFINED AN ORE BODY AS SUCH. HOWEVER, THE GEOLOGICAL MAPPING AND NUMEROUS SHOWINGS GIVE US EVERY CONFIDENCE, THAT WITH EXTRA WORK, WE WOULD BE ABLE TO DO THAT. SIMILAR DEPOSITS IN FRANCE LEAD US TO BELIEVE THAT ONCE WE HAVE FOUND ONE DEPOSIT WE WILL FIND A LARGE NUMBER OF SMALL SIMILAR DEPOSITS THROUGHOUT THIS AREA.

THERE ARE OTHER ADVANTAGES TO THIS PARTICULAR LOCATION OTHER THAN EASY DRILLING AND EXTRACTION. WINDSOR IS THE NEAREST TOWN OF ANY SIZE. IT HAS A FULL INTRASTRUCTURE OF SHOPS, SERVICES, SCHOOLS, POLICE FORCE, ETC. TRANSPORT ACCESS IS VERY GOOD. THE HIGHWAY NO. 14 RUNS TO WITHIN 5 OR 6 MILES OF THE SITE. A RAIL-LINE RUNS THROUGH THE TOWN OF WINDSOR. WE ARE LESS THAN ONE HOUR'S DRIVE FROM THE PORT CITY OF HALIFAX. CURRENTLY, THERE ARE OVER 2000 UNEMPLOYED IN THE AREA, AND MANY OF THESE PEOPLE HAVE TRADE OR SIMILAR SKILLS.

AS I SAID, WE CONTINUED DRILLING UNTIL SEPTEMBER, 1981 WHEN THE PROVINCIAL GOVERNMENT ISSUED A MORATORIUM ON URANIUM LICENSES. THIS TOOK PLACE SHORTLY BEFORE THE PROVINCIAL ELECTION. IN JANUARY THIS YEAR, THE NOVA SCOTIA GOVERNMENT APPOINTED JUDGE ROBERT MCCLEAVE TO CONDUCT A PUBLIC INQUIRY INTO THE IMPACT OF URANIUM MINING, MILLING AND EXPLORATION. DUE THE THE MORATORIUM, WE ARE DOING NO EXPLORATION WORK AT MILLET BROOK. WE ARE, HOWEVER, CONTINUING THE ENVIRONMENTAL MONITORING

OF THE AREAS IN PREPARATION FOR AN ENVIRONMENTAL IMPACT STUDY. THIS IS A CONTINUATION OF AN EXISTING TWO YEAR PROGRAM.

THE PUBLIC INQUIRY IS BEING HELD UNDER THE PUBLIC INQUIRIES ACT. TO DATE JUDGE MCCLEAVE HAS RECEIVED 270 APPLICATIONS TO MAKE PRESENTATIONS BEFORE HIM. THESE WILL BE HEARD DURING THE FIRST STAGE OF THE PUBLIC INQUIRY, WHICH THE JUDGE CALLS THE PRESENTATION STAGE. HE IS HOLDING THESE HEARINGS IN TOWNS THROUGHOUT THE PROVINCE. JUDGE MCCLEAVE HAS OBVIOUSLY MADE GREAT EFFORT TO MAKE THESE HEARINGS ACCESSIBLE TO EVERY PERSON WHO WISHES TO MAKE A PRESENTATION. I SUSPECT IN DOING SO HE HAS CREATED MANY LOGISTICAL PROBLEMS FOR HIMSELF. HOWEVER, HE HAS ALSO MADE IT QUITE CLEAR THAT HE WISHES FOR A VERY ORDERLY MEETING.

THE PRIMARY ISSUES BEING RAISED BY CONCERNED CITIZENS AND OPPOSITION GROUPS I AM SURE ARE FAMILIAR TO YOU ALL. I WAS READING THROUGH THE LIST OF OBJECTIONS GIVEN TO MR. JUSTICE PARKER AT THE WINDSCALE INQUIRY IN 1977. WITH A FEW MINOR CHANGES THEY REPRESENT THE SAME OBJECTIONS BEING RAISED IN NOVA SCOTIA. FOR THE SAKE OF COMPLETION, I WILL LIST THEM HERE:

1. HEALTH HAZARDS - BOTH TO THE GENERAL PUBLIC AND THE MINE EMPLOYEES - INCLUDING, AND IN PARTICULAR, LUNG CANCER.
2. THE RISK AND HAZARDS OF RADIAITON - IN PARTICULAR THE LOW-LEVEL RADIATION CONTROVERSY.
3. THE EFFECT OF RADIATION ON GENETICS AND THE RESULTANT MUTATIONS. MUCH IS MADE OF THE KERALA STUDY.

4. TAILINGS AND WASTE MANAGEMENT PROBLEMS - THIS IS ONE OF THE MAJOR POINTS OF OBJECTION. ELLIOT LAKE IS FREQUENTLY USED AS AN EXAMPLE. SOME OF THE DESCRIPTIONS OF ELLIOT LAKE ARE EXAGGERATED TO THE POINT WHERE THEY ARE LUDICROUSLY FUNNY.
5. THE INVOLVEMENT BETWEEN URANIUM MINING, NUCLEAR POWER AND NUCLEAR WEAPONS
6. THE CURRENT STATE OF THE INDUSTRY - ESPECIALLY ITS ECONOMIC STATE. THE EFFECT OF THE URANIUM MINE ON TOURISM, FISHING, FORESTRY AND FARMING.

OVER THE LAST SEVERAL YEARS, THERE HAVE BEEN A NUMBER OF IMMIGRANTS TO NOVA SCOTIA FROM THE NORTHEASTERN UNITED STATES, PARTICULARY THE INDUSTRIALIZED AREAS. A LOT OF THESE IMMIGRANTES HAVE FAIRLY ADVANCED TECHNICAL TRAINING. (I.E. SCHOOL TEACHERS, JOURNALISTS, RESEARCHERS, ETC.) THE REASON FOR THIS INFLUX APPEARS TO BE A STRONG DESIRE TO LEAVE THE INDUSTRIAL AREAS BEHIND. MANY OF THEM APPEAR TO BE AFFLUENT ENOUGH TO PURCHASE SMALL FARMS. HERE THEY HAVE MADE A MAJOR EFFORT AS SELF-SUBSISTENT AS POSSIBLE. RAISING THEIR OWN FOOD, EGGS, CHICKENS, GOATS AND USING ALTERNATE FORMS OF ENERGY LIKE WOOD STOVES. IN SOME CASES, THIS HAS BEEN TAKEN TO THE EXTREME, IN THAT PEOPLE ARE USING LOG CABINS WITH NO RUNNING WATER OR ELECTRIC LIGHT. MOST OF THESE PEOPLE HAVE COME WITH STRONG ANTI-TECHNOLOGY AND IN PARTICULAR STRONG ANTI-NUCLEAR POWER EMOTIONS.

THEY LOOKED UPON NOVA SCOTIA AS AN UNSPOILED PROVINCE WHERE THEY CAN RETREAT FROM THE HORRORS OF CIVILIZATION. I'AM SURE IT MUST HAVE BEEN A MAJOR BLOW FOR THEM WHEN THEY LEARNED OF THE DECISION TO BUILD POINT

LEPREAU AND AN ANTI-NUCLEAR POWER REACTOR GROUP FORMED CALLED F.A.C.T. (FUNDY AREA CONCERNED FOR TOMORROW). THIS GROUP HAS SHOWED THE VERY EMOTIONAL HELEN CALDICOTT TAPE IN MANY PLACES. THEY HAVE ALSO EXPRESSED GREAT CONCERN ABOUT THE EFFECTS OF POINT LEPREAU ON THEIR ENVIRONMENT. I WOULD ADD THAT POINT LEPREAU IS 50 MILES AWAY FROM WHERE MOST OF THESE PEOPLE LIVE.

AS YOU CAN SEE THERE WAS ALREADY AN ANTI-NUCLEAR MOVEMENT IN PLACE BEFORE OUR PEOPLE DISCOVERED URANIUM. IT MUST HAVE BEEN AN EVEN MORE DISAPPOINTING SHOCK TO DISCOVER THAT PRISTINE NOVA SCOTIA CONTAINED DEPOSITS OF URANIUM. I THINK AT THE BEGINNING, OUR GEOLOGISTS WERE AMAZED AT THE SPEED AND THE MAGNITUDE OF THE RESPONSE THAT WAS STIRRED UP BY THEIR DISCOVERIES. MANY OF THEM WERE TAKEN ABACK AND OVERWHELMED BY THE VOLUME OF SO CALLED FACTS WHICH WERE SPREAD TO THE POPULACE. NOW THAT THEY HAVE EXPERIENCED THIS FOR A YEAR OR SO, THEY REALIZE REASONS FOR THESE REACTIONS. WE NOW HAVE A LARGE NUMBER OF ANTI-NUCLEAR GROUPS ALL WITH THEIR OWN ACRONYM C.A.P.E., C.A.U.M., R.E.S.C.U.E. AND SO ON. ANOTHER VERY PROMINANT GROUP IS THE ECOLOGY ACTION CENTRE. THIS GROUP LEADS MANY CITIZENS GROUP ACTIONS AGAINST A VARIETY OF INDUSTRIAL INDEAVORS. FOR EXAMPLE, THEY HAVE BEEN VERY ACTIVE IN THE FIGHT AGAINST SPRUCE BUDWORM SPRAYING. IN FACT, THEY HAVE BEEN SUCCESSFUL IN PREVENTING SPRUCE BUDWORM SPRAYING IN NOVA SCOTIA.

WHEN I ARRIVED IN THE PROVINCE LAST YEAR, THERE WERE A LARGE NUMBER OF PAMPHLETS BEING PRODUCED AND CIRCULATED, MANY OF WHICH MADE VERY WILD AND EXAGGERATED CLAIMS ON THE URANIUM ISSUE. THESE HAVE CREATED A GREAT DEAL OF CONCERN AMONGST THE LOCAL PEOPLE AND THE ANTI-NUCLEAR GROUPS HAVE BEEN SUCCESSFUL IN GETTING A NUMBER OF PROMINENT GROUPS (THE WOMEN'S

INSTITUTE, FEDERATION OF AGRICULTURE, VALLEY MEDICAL ASSOCIATION) TO ISSUE A CALL FOR A MORATORIUM. WHEN MANY OF THESE PEOPLE APPEAR IN FRONT OF THE INQUIRY THEY EXPLAIN TO THE JUDGE THEIR LACK OF SCIENTIFIC BACKGROUND AND STATE QUITE REASONABLY THAT AS THEY HAVE ONLY SEEN EVIDENCE TO SAY URANIUM WILL BE A BAD THING FOR NOVA SCOTIA, WE MUST ASK FOR A MORATORIUM. SOME OF THEM ADD THAT IF EVIDENCE COULD PROVE OTHERWISE, THEY WOULD DROP THEIR OBJECTIONS.

I WOULD LIKE TO MAKE ONE OTHER COMMENT AND BASED ON MY EXPERIENCE IN ONTARIO, IT'S A VERY PERTINENT ONE. IN GENERAL, WE HAVE RECEIVED VERY FAIR TREATMENT FROM THE NOVA SCOTIA MEDIA. THIS INCLUDES RADIO, TELEVISION, AND THE NEWSPAPERS. I DO NOT MEAN TO IMPLY BY THIS THAT THE MEDIA ARE BIASED IN OUR FAVOR BUT SIMPLY THAT THEY HAVE NOT TRIED TO DISTORT ANY OF OUR STATEMENTS OR TO TRY TO MAKE US LOOK FOOLISH. BECAUSE OF THIS, WE HAVE TRIED TO BE AS HELPFUL AS POSSIBLE AND HAVE WILLINGLY GIVEN INTERVIEWS AND STATEMENTS. IN FACT, WE HAVE HAD ONLY ONE CASE WHERE I WAS VERY DISPLEASED WITH THE DISTORTION AND INNUENDO PLACED ON STATEMENTS THAT I MADE.

AT THIS DATE, THE INQUIRY IN ONLY PARTWAY THROUGH. WE HAVE MANY MORE HEARINGS TO COME INCLUDING OUR OWN COMPANY'S PRESENTATION. THE NEXT STAGE OF THE INQUIRY WILL BE WHAT JUDGE MCCLEAVE CALLS THE CONFRONTATION STAGE WHERE EXAMINATION AND CROSS-EXAMINATION WILL TAKE PLACE. THIS WILL OCCUR IN THE EARLY FALL. THIS WILL BE FOLLOWED BY A SUMMATION BY THOSE PARTIES TAKING PART IN THE SECOND STAGE. THE FINAL REPORT IS DUE BY THE END OF THE YEAR.

SLIDES

N.S. GEOLOGICAL MAP

MILLET BROOK GELOGICAL MAP

AERIAL PHOTOS - MILLET BROOK (1)
(2)
(3)

ROCK SAMPLES

DRILLING RIG

BERNADAN SLIDES (3)

DISCUSSION OF PAPER BY R. JOHN

R. Laelato	What is the anticipated cut-off grade of ore?
R. John	That decision has not been made, but it may be 0.2-0.4%. The ore appears easy to process. There will be at least 2 years exploration required before the feasibility study is made.

R&D ACTIVITIES AT ELDORADO NUCLEAR LIMITED - AN OVERVIEW

I. Itzkovitch

Eldorado Nuclear Ltd.

NO PAPER AVAILABLE

REPORT OF IAEA AND FORT COLLINS MEETINGS

M.C. Campbell
CANMET

REPORT OF IAEA AND FORT COLLINS MEETINGS

AS YOU KNOW I AM THE CANADIAN DELEGATE TO THE NEA/IAEA WORKING GROUP ON URANIUM EXTRACTION. IN ADDITION, I AM THE CURRENT CHAIRMAN OF THAT GROUP. AS SUCH I HAVE ATTENDED TWO MEETINGS IN THE PAST YEAR WHICH HOLD SOME INTEREST FOR CUPMC MEMBERS. IN FACT, I LOOK ON THE NEA/IAEA WORKING GROUP ON URANIUM EXTRACTION AS THE INTERNATIONAL EQUIVALENT OF THE CUPMC, AND I WOULD LIKE TO USE THE CUPMC AS A FORUM THROUGH WHICH TO MAINTAIN COMMUNICATIONS BETWEEN THE CANADIAN INDUSTRY AND THE WORKING GROUP. INCIDENTALLY, THE WORKING GROUP IS NOT A CLOSED SHOP, YOU ARE WELCOME TO ITS MEETINGS AND TO PARTICIPATE IN ITS ACTIVITIES, HOWEVER EACH ATTENDEE MUST BE NOMINATED BY ITS NATIONAL GOVERNMENT, MORE A FORMALITY THAN A CONSTRAINT.

I ATTENDED TWO MEETINGS:

NEA/IAEA WORKSHOP ON URANIUM ORE PROCESSING TAILINGS CONDITIONING FOR MINIMISING ENVIRONMENTAL PROBLEMS IN TAILINGS DISPOSAL, OCTOBER 28-29, 1981 IN FORT COLLINS, COLORADO

5TH MEETING OF THE NEA/IAEA WORKING GROUP ON URANIUM EXTRACTION, FEBRUARY 25-27, 1982 IN VIENNA, AUSTRIA.

1. THE PROCEEDINGS OF THE FORT COLLINS WORKSHOP WERE RECENTLY PUBLISHED BY THE NEA AND ARE AVAILABLE FROM THE NEA. I CAN GIVE YOU A REFERENCE. A PARALLEL WORKSHOP WAS HELD ON GEOMORPHOLOGICAL ASPECTS OF TAILINGS.

URANIUM MILL TAILINGS MANAGEMENT

PROCEEDINGS OF TWO NEA WORKSHOPS ON:

GEOMORPHOLOGICAL EVALUATION OF THE LONG-TERM STABILITY OF URANIUM MILL TAILINGS DISPOSAL SITES

AND

URANIUM ORE PROCESSING/TAILINGS CONDITIONING FOR MINIMIZING LONG-TERM ENVIRONMENTAL PROBLEMS IN TAILINGS DISPOSAL

COLORADO STATE UNIVERSITY
FORT COLLINS

OCTOBER 28-30, 1981

AVAILABLE FROM: DIRECTOR OF INFORMATION, OECD
2 RUE ANDRE-PASCAL
75775 PARIS CEDEX 16, FRANCE

THE MEETING PROVIDED A FORUM FOR THE REVIEW AND DISCUSSION OF VARIOUS METALLURGICAL OPTIONS WHICH COULD BE USED TO REDUCE THE LEVEL OF POLLUTANTS DISCHARGED BY URANIUM MILLS TO TAILINGS. ESSENTIALLY THREE OPTIONS WERE PRESENTED:

OPTION 1:

USE TAILINGS DISCHARGE AND DISPOSAL TECHNIQUES WHICH WOULD ELIMINATE SEEPAGE. DR. ROBINSKY OF THE UNIVERSITY OF TORONTO DESCRIBED THE THICKENED DISCHARGE TECHNIQUE. THIS METHOD WAS DESCRIBED AS TECHNICALLY EFFECTIVE AND AT A COST COMPARABLE TO CONVENTIONAL PONDING.

OPTION 2:

PHYSICAL SEGREGATION OF ORE OR TAILINGS INTO FRACTIONS USING WELL KNOWN MINERAL DRESSING TECHNOLOGY. DR. SIMONSEN OF SOUTH AFRICA DESCRIBED PRECONCENTRATION TECHNOLOGY. DR. LEVINS OF AUSTRALIA REVIEWED THE MERITS OF SPLITTING TAILINGS INTO DIFFERENT FRACTIONS TO ISOLATE THE RADIONUCLIDES AND ACID PRODUCING FRACTIONS. DR. DREESON OF THE LAS ALAMOS NATIONAL LABORATORY DESCRIBED THERMAL STABILIZATION TECHNIQUES FOR SAFE DISPOSAL OF RADIONUCLIDES. COSTS OF THESE SEGREGATION TECHNIQUES WERE ESTIMATED TO BE IN THE ORDER OF $1.00 A TON OF TAILINGS WITH STABILIZATION COSTS OF UP TO $20. A TON OF THE SEGREGATED FRACTIONS.

OPTION 3:

THIS INVOLVED THE CONCENTRATION AND ISOLATION OF CONTAMINANTS BY AGGRESSIVE CHEMICAL MEANS. MR. AL RYON OF OAK RIDGE NATIONAL LABORATORY SUMMARIZED WORK ON NITRIC ACID LEACHING. I (MCC) REVIEWED WORK AT CANMET ON CHLORIDE METALLURGY. COSTS WERE ESTIMATED TO BE $5 TO $10 PER POUND URANIUM HIGHER THAN CONVENTIONAL TECHNOLOGY.

SEVERAL OTHER PAPERS PROVIDED INSIGHTS INTO RELATED WORK ON THE MOBILITY OF SALTS IN TAILINGS (DR. MARKOS) AND LEACHING STUDIES (DR. BAIRD, McMASTER UNIVERSITY, DR. BLAND, UNIVERSITY OF CALGARY, AND MR. LANDA OF THE U.S. GEOLOGICAL SURVEY).

THE WORKSHOP CONCLUDED THAT SITE SPECIFIC FACTORS WERE IMPORTANT, THAT CONTAMINANTS OTHER THAN RADIUM MUST BE CONSIDERED, AND THAT GOOD PREVENTATIVE TECHNOLOGY WAS CERTAINLY CHEAPER THAN REMEDIAL ACTION. REPORTS BY THE US. DOE, EPA AND NRC INDICATED TO A JOINT SESSION PRIOR TO THE WORKSHOP THAT REMEDIAL ACTION ON 24 TAILINGS SITES CONTAINING 25 MILLION TONS OF TAILINGS WOULD COST $380 MILLION. THE URANIUM EXTRACTION WORKSHOP REQUESTED OF THE NEA/IAEA TAILINGS WORKING GROUPS THAT GUIDANCE BE PROVIDED ON THE NEED TO DEVELOP AGGRESSIVE CONTAMINANT ISOLATION TECHNOLOGY TO MEET THE ENVIRONMENTAL REGULATIONS.

2. THE 5TH MEETING NEA/IAEA WORKING GROUP ON URANIUM EXTRACTION WAS WELCOMED BY MR. SYED FAREEDUDDIN, DIRECTOR OF THE DIVISION ON NUCLEAR FUELS AT THE IAEA IN VIENNA. MR. FAREEDUDDIN SPENT SOME TIME AT THE MINES BRANCH IN THE EARLY 1950'S.

EACH ATTENDEE PRESENTED A SHORT OVERVIEW OF RESEARCH ACTIVITIES IN THEIR COUNTRY. THESE REVIEWS WILL BE SUMMARIZED IN A NEWSLETTER TO BE PUBLISHED BY THE NEA SHORTLY. I WILL TRY TO HAVE COPIES OF THE NEWSLETTER SENT TO REPRESENTATIVES OF EACH CANADIAN COMPANY. IF YOU RECEIVE ONE PLEASE CIRCULATE IN YOUR ORGANIZATION.

MUCH OF THE MEETING WAS SPENT DISCUSSING A PUBLICATION ENTITLED "URANIUM EXTRACTION TECHNOLOGY: CURRENT PRACTICE AND NEW DEVELOPMENTS IN ORE PROCESSING" TO BE ISSUED BY THE NEA IN SEPTEMBER 1982. THIS BOOK WILL HAVE FIVE SECTIONS:

INTRODUCTION
URANIUM RESOURCES: GEOLOGY AND EXPLOITATION OF RESOURCES
ADVANCES IN TECHNOLOGY
ECONOMIC ASSESSMENT
COUNTRY REPORTS

IT IS INTENDED TO PROVIDE A USEFUL AND PRACTICAL GUIDE TO THOSE NOT FULLY CONVERSANT WITH THE EXTRACTIVE METALLURGY OF URANIUM. THERE IS A PARTICULAR NEED FOR SUCH A DOCUMENT IN THIRD WORLD COUNTRIES SERVICED BY THE IAEA WHERE TOO OFTEN RESOURCES ARE EQUATED ONLY IN TERMS OF MARKET PRICE FOR YELLOW CAKE RATHER THAN THE REAL COST OF PRODUCTION.

THE NEXT WORKING GROUP MEETING WILL BE HELD IN PARIS IN APRIL 1983 IN CONJUNCTION WITH A WORKSHOP ON METHODOLOGY FOR COST ESTIMATION.

IN LATE 1983 THE IAEA WILL SPONSOR A MEETING IN VIENNA ON ADVANCES IN URANIUM ORE PROCESSING AND RECOVERY FROM NON-CONVENTIONAL RESOURCES. IF YOU WISH TO PARTICIPATE IN THESE PLEASE CONTACT ME. IN ANY CASE, I WILL KEEP YOU ABREAST OF DEVELOPMENTS AS I AM INFORMED.

URANIUM AND RARE EARTHS IN CHINA

G.M. Ritcey
CANMET

URANIUM AND RARE EARTHS IN CHINA

G.M. Ritcey
CANMET

INTRODUCTION

During the month of October 1981, I was fortunate in visiting China. The visit to China was at the joint invitation of the ACADEMIA SINICA, Institute of Chemical Metallurgy in Peking and the Institute of Organic Chemistry in Shanghai. The Chinese are bringing in many "experts" from all fields to speak and lecture to their scientists in order to bring them up to date in the various scientific disciplines. Their Cultural Revolution of 1966-1976 resulted in considerable damage to the universities and institutes and much of their data was destroyed. Scientists were sent periodically to the farms to work, or to work on the railway, etc. during the period. They are quickly regaining what they lost and their present technology is impressive in many areas of metallurgy.

In China there are 117 institutes under the ACADEMIA SINICA with a total of about 76,000 people. Many of the institutes I visited had 800-1000. The institute also is responsible for housing and schooling for children, therefore the 1000 personnel includes the manpower necessary for the total running and maintenance of all facilities. The Chinese were most kind and considerate throughout the visit. They provided a scientist interpreter guide for the visit to each area, vehicle, and arranged all visits, travel and hotels.

The scientists, wherever I visited, were technically well informed- and obviously were up to data on the literature. Most of their equipment is designed and manufactured in China, including computer control facilities.

I found the visit at each establishment most informative, with the personnel providing answers to any qeustions asked. Their level of expertise was at times impressive, and they appeared well organized with respect to co-operative research between institutes, universities and industry.

At this meeting of CUPMC I will briefly describe aspects of their progress in uranium and rare earths.

URANIUM

Visits were made to the Uranium Institute and to a mill in Henyang. The level of research at the Institute was excellent and the mill was one of the cleanest and with the most concern for the worker protection of all mills visited anywhere.

Uranium Institute, Beijing

The uranium institute was established in 1958 and is located 50 km outside the City of Beijing. Of the 1000 personnel, 500 are engineering and technical. The research is divided into Ore Dressing, Hydrometallurgy, Analytical, Effluents Treatment and Organic Synthesis.

Ore Dressing

This section is composed of radiometric sorting, flotation, magnetic separation, gravity concentration, high intensity magnetic separation and mineralogy. They do considerable investigations with electromagnetic fluid, have an electromagnetic separator from the UK (much of their equipment is built on-site), and a high frequency dielectric separator. In radiometric sorting they have an excellent unit which measures the activity of the uranium ores by counting the gamma, displays the amount of uranium in the ore (in grams or kilograms) as well as the grade of ore. This appeared to be an excellent unit which was made at the Institute and which only takes about 20 seconds to obtain the complete analyses. The unit was so designed as to take samples from a few grams up to several kilograms in weight. Some work has been done on carbonate flotation to float off the acid consumers.

Leaching

Pressure alkaline leaching is done on carbonate ores; while for sulphide ores an autoclave leach with oxidation to produce H_2SO_4 is used as the leachant. For refractory ores they have been using a hot pug H_2SO_4 leach. For low grade ores they have investigated bacteria leaching.

In the leaching lab itself, tests were underway using a silicate ore. They are encountering chlorides which are acid consumers and which consume more acid as the leach temperature is increased. Tests are at a constant pH in order to obtain less impurity dissolution. The leach conditions are pH 2.0, at 45°C for 4 hours in a constant temperature bath using magnetic stirrers. They have a series of these set up so they can do about 6 leaches simultaneously. Normal acid requirements are 30 kg acid/ton ore, which rises to 100 kg acid/ton ore as the temperature is increased to 80°C from ambient temperature.

They are doing pressure leaching on sulphide ores containing 3% pyrite. These are fines from the H_2O wash elutriation described below. The leach is under pressure on the H_2O slurry, 140°C, 3.5 hours in a stainless steel container at 14% solids.

By product recovery is a concern, and from some ores they recover Cu, Pb, Zn.

Solid-Liquid Separation

The laboratory is well equipped, and they recognize that it is important to investigate the particle size as well as the addition of flocculents. Much work has been done on flocculents, and investigation of thickeners, such as the Enviroclear plate clarifiers. They have been investigating the use of columns for washing of solids, by using a fluidization method. Flocculent is added about 1 foot from the top of the column (in their lab demonstration). The water to fluidize the solids is added at the bottom of the column. The interface level is maintained about half way up the column by an ultrasonic device. A star valve is being used to discharge the effluent. They find the column procedure much more effective than thickeners, because of (a) decreased space requirement; (b) the underflow is much more dense; (c) the concentration of uranium in the overflow is higher; and (d) there is less uranium soluble loss. For example in the conventional method, there is 300 ppm U in the overflow. By using the column, this can be upgraded to about 1 g/L. Such a procedure works effectively in alkaline leach pulps, but is not as successful with acid pulps. It can be used for classification and sand/slimes separation.

Purification

They use ion exchange or solvent extraction and have investigated resin in pulp (RIP) and are presently looking at solvent in pulp (SIP). The RIP employs the fluidized bed technique, and they are up to 8 wt % solids. In the ion exchange laboratory they had a number of very small columns set up to examine different resins for chemical and physical properties. Equilibrium testing was in progress on quaternary and tertiary resins.

In the solvent extraction of uranium they are involved in the synthesis of extraction reagents and the investigation of SIP where they are using 20% solids with an amine. They are using 2-inch diameter pulse columns and mixer settlers. Work on the uranium extraction from coal lignite ash includes the recovery of germanium and molybdenum as well as uranium. Uranium is recovered from phosphoric acid using DEHPA-TOPO mixture. The phosphoric acid process is a one cycle process, instead of the conventional 2-cycle route, where they produce ammonium uranyl tricarbonate (AUTC) as their stripped product.

The uranium work with quaternary amines is from carbonate solutions, where they have a small set-up of mixer settlers. They are concerned with anion impurities (NO_3^-) as well as humic acids. Nitrate poisoning is being examined. They have on-line analyses for uranium and for control.

With respect to SIP, they have carefully repeated and confirmed all of the CANMET work on the subject. They agree that SIP is the major effort that should be directed in hydrometallurgy research for the cost savings to the operation are significant. I was most impressed with the development work in this area. They had looked at pulse columns and are now working on mixer settlers of their own design, with inclined plates to transfer the solids from stage to stage. They have been recovering uranium from the treatment of lignite, using 4 stages of extraction and 1 stage for separation prior to tailings discharge. The amine loss in such a pulp circuit is 150 g amine/tonne lignite ash (1% U in ash). This material contains high Si and fines which make it difficult to filter, and therefore SIP is an attractive route. They also have a pulse column set up and will continue work on that contactor starting where we terminated the CANMET work of several years ago.

With respect to stripping uranium from amine circuits they are in favour of using something other than NH_3-$(NH_4)_2SO_4$ which is currently used in most mills. They are producing AUTC after stripping with $(NH_4)_2CO_3$. They would like to see the AUTC product produced, but all buyers want "yellowcake". They also produce UO_2, UO_3, U_3O_8 and UF_4. As by-products from uranium processing, they produce Mo_3O_4 and ammonium and potassium paramolybdates.

Organic Synthesis

This section develops and manufactures flocculants, ion exchange resins, solvent extraction reagents, and analytical reagents to match with the processing requirements. In ion exchange, they have heavy resins approaching a density of 1.31 which they are using in fluidized continuous ion exchange. They have investigated the production of quaternary amines as well as other reagents for solvent extraction. They appear to have phosphonic and phosphinic acids.

Analytical

Well set up laboratories and much equipment were in evidence. They had an inductance coupled plasma torch for the determination of impurities in waste water down to ppb range. They analyse 10 elements simultaneously and use a 12-channel direct reading emission spectrometer for the determination of micro quantities of constituents. Their X-ray fluorescence was a Japanese make, fully automatic with data print out, 6 samples simultaneously, and could analyse 200 samples per shift. Also they are successful in analysing directly loaded resins for Fe, Mo, Si, P, S, U. The sample of resin is placed into the unit without any prior sample preparation. Also they have been analysing loaded solvents without any prior treatment.

They have a micro diffractometer, 12 Kw, requiring 2 hours per sample. They can differentiate between U_3O_8 and UO_2 but not between U^{+4} and U^{+6}.

Environment

All effluents must be treated. They are concerned with air as well as the soil and liquid effluents. Barite is used for radium co-precipitation removal from tailings effluents to achieve 3×10^{-11} curies/gram. Coagulation precipitation is used for removal of fluoride from waste waters. Electrodialyses is used to remove NaCl from effluents and the H_2O is recycled to the process.

Public acceptance of nuclear energy is a concern in China as in other parts of the world. There is still a reaction by the public to the use of radioactive materials.

Uranium Ore Processing Plant, Hengyang, Hunan Province

The plant is located in the southern part of China, in Hunan Province, south of Changsha, and accessible by rail from Shanghai, Changsha or Canton.

The mill was constructed in 1958 and put into operation in 1962. There are about 4000 on the property, which is really a town site. There are 1200 workers and a staff of 200 for the operation, which is on 520 hectares.

Feed to the plant is from a nearby mine and grades up to 0.1% U_3O_8. The original design of the plant was for 800,000 tonnes/year and the products are 1000 tonnes UO_2/year. It is a plant that goes through the production of ADU and refining to UO_2. The feed is a complex carbonaceous sandstone with granite being the host rock. (Other ore types have apparently been treated, containing brannerite and pitchblende, but they were processing sandstone at the time of visit). The leaching efficiency is 91%, while the total recovery throughout the plant is 86-87%.

Ore Preparation

The ore is delivered by conveyor belt to the coarse ore bin. They employ gamma-beta measuremants on the belt. The ore is reduced to 300 mm by jaw crushing, and to 150 mm with a cone crusher and sent to a vibrating screen. The oversize is recycled to another crusher. The undersize material from both crushers is sent to the vibrating screen. The undersize from the screen is sent to the ball mill (2.7 m diameter x 2.7 m long) which is in closed circuit with a spiral classifier.

Leaching

Pulp from the grinding section is sent to the hydrocyclone (250 mm diameter). The underflow goes to the pachuca leach towers. The overflow goes to the 2nd of 4 towers. The H_2SO_4 is added to the 1st tower. The leaching is at 75°C, 50% solids for 5 hours. In the final hour, the acidity is controlled to 40-45 g H_2SO_4/L. They have 4 banks of 5 pachucas in each bank. Some of the ores require an oxidant, and pyroluscite has been used. They prefer not to use this because of the manganese impurity and were therefore interested in CANMET's work on oxidants.

Sand/Slimes Separation

The leach slurry is sent to sand/slimes separation and washing. The hydrocyclones are in closed circuit with the classifier (5 classifiers, 2 for separation and the others for washing). The wash water is added only in the final stage.

The classifier overflow is pumped to the hydrocyclone. After the 2-stage wash, the pulp is sent to a roto screen to remove any wood chips, and from there to neutralization. The underflow from the hydrocyclone is sent to another hydrocyclone section. The tailings are neutralized to pH 9-10 with lime slurry. No $BaCl_2$ is added. The levels they try to achieve are 5×10^{-11} pCi/L Ra^{226}.

The leach solution is adjusted to pH 2.0 by means of lime slurry. Wash H_2O is added to a pulp density of 8-10%.

Resin-in-Pulp (RIP) Ion Exchange

The 8-10% slurry is fed directly to the RIP circuit. The circuit consists of 3 banks of 5 stages of adsorption, 2 for regeneration, and 1 for elution. Each column is 3.5 meters diameter and 20 meters high. They use a strong base resin, similar to the IRA-400. The slurry is introduced into the bottom of the extraction column. When the breakthrough reaches 5 mg/L, then the feed is loaded to the next column. After loading, the resin is flushed with H_2O. They have used 3.5% NH_4NO_3 + NH_4OH (5%) as the eluant. A H_2O flush removes any NH_4^+. (They have also used HNO_3 for elution). The eluate contains 10 g U_3O_8/L. Resin losses were stated as 7 grams resin per tonne of feed.

Refining

The eluate is sent to the purification or refinery section of the plant. This is a solvent extraction circuit, using 18% TBP in kerosene, using two sets of packed bed (rachig rings) columns in series. One stage was used for each of extraction, scrubbing, stripping, washing and solvent clean-up. These columns are 4 feet diameter and 30 feet high for extraction, and 4 feet diameter and 24 feet high for the stripping and scrub columns. At one time they tried sieve plate pulse columns, using air pulsation, but were not completely satisfied.

Following scrubbing, the stripping is carried out at 60°C using slightly acidified water (1-2 g/L HNO_3). The uranium nitrate is pumped to the precipitation section, consisting of 4 agitator tanks in series, where the pH is controlled at pH 2.5 in the first stage and pH 6.5 in the second stage, using ammonia. They produce an ADU intermediate product, which is thickened in 2 stages and settled, then filtered in plate-and-frame filters. The product is then slurried in 500 g/L $(NH_4)_2CO_3$ to convert the ADU to ammonium uranyl tricarbonate (AUTC), which is sent to hydrocyclones and then to clarification. The underflow from the hydrocyclones goes to disc filters. The AUTC contains 8% H_2O, and this cake is roasted at 880-900°C in parallel horizontal roasters, 7.8 meters long and 450 mm diameter. The reduction takes place in the last section of the furnace. The UO_2 product is cooled and packaged.

In addition to producing UO_2 as a product, they also calcine to a U_3O_8 product and of course the ADU if there is a demand for it. The AUTC is converted to UF_4 in another plant in China as an intermediate to fuel production. No satisfactory answer was given regarding the number of uranium plants in China.

They have possible by-products of La and Mo in the circuit, but at present they find it too costly to recover them and therefore they have done nothing about it.

Most of the equipment throughout the plant is stainless steel. They therefore do not wish to change from pyrolucite to $NaClO_3$. Therefore they are interested in other methods of oxidation.

The plant shuts down twice a year for a 2-week maintenance. The housekeeping throughout was excellent and the plant very clean.

RARE EARTHS

China has enormous deposits of monazite and bastnasite, much more than the Molycorp bastnasite deposits in Mountain Pass, California, which have always been stated to be the world's largest deposit. There are also deposits of xenotime and rare earths occurring as by-products in iron ore. The proven reserves of mixed rare earths in China are 36 million tonnes compared to Molycorp's Mountain Pass reserves of 3 million tonnes. They can process 1500 tonnes monazite yearly, and vary production to meet the annual requirements.

They are building a new facility for Y production. The building has just been completed and some personnel and facilities are now there. Yttrium accounts for 2-3% of the total rare earths in the monazite. The present price of Y is $80 US/kilo (1982).

Rare Earth Institute, Baotou, Inner Mongolia

The Rare Earth Research Institute at Baotou was set up in 1963 because of the rare earths in the region, particularly that contained in their iron ores. During the Cultural Revolution between 1966-1976 the Institute was damaged and many records lost.

The Institute is composed of 800 people, with 300 professionals, 300 workers and 60 adminstration. The remainder service the buildings, living quarters, teach school, look after children, etc. There are more than 100 in the workshops, responsible for the maintenance of research equipment and the building of equipment.

The following summarizes the various areas:

Pre-Treatment and Flotation - 80 people

Hydrometallurgy - 40 people, 20 being professional. They are engaged in solvent extraction and rare earth separations.

Pyrometallurgy - Again about 40 people, 20 being professionals in addition to 20 technicians and maintenance.

Niobium Group - There is a special group of 30 people devoted to niobium because the nearby mine contains much niobium.

Steel Materials - There are more than 80 people working mainly on the rare earth alloys, their behaviour and use.

Physical Examination - There are 40 people engaged in examining the various properties of alloys and the other various materials they make, and also they research the behaviour of rare earths in steels, such as tensil strength.

Analytical Chemistry - 80 people.

Instrument Department - There are 30 people engaged in the maintenance and in the building of equipment.

Technical Information Department - This is service to research to provide technical information when required (similar to CANMET TID).

New Materials in Rare Earths - This group is 40 people, involved in magnetic materials, particularly the permanent magnets of Sm-Co.

Workshop - There are 30 in this area, responsible for the production of permanent magnets, also they work on energy storage, the investigation of electronegative materials, and research applications in other areas.

Hydrometallurgy

The hydrometallurgy research at the Institute is based mainly on the needs of the nearby Baotou mine, and involves leaching, solvent extraction and recovery. There are three methods of leaching:

1. NaOH leach, followed by HCl leach of the residue to yield 90%.
2. Na_2CO_3 roast, followed by H_2SO_4 leaching to yield 85%.
3. H_2SO_4 leaching of the concentrate, but only in laboratory size at this time.

For the separation of the rare earths, they formerly used ion exchange, but now it is by solvent extraction. The ion exchange circuits employed EDTA to enhance the separation (after work of Ames Labs and probably similar to Michigan Chemicals). In the present leaching procedure the acid is too strong and so would affect the resin. By using IX, they obtained:

99 to 99.9% Pr
99 to 99.99% Nd
99 to 99.9% Sm
99 to 99.99% Gd

They are using DEHPA for SX, also a neutral phosphate and amines. The separation of rare earths from impurities and other rare earths is the object. The main impurities are Fe, P, Th. The rare earths in the concentrate contain mainly the light and medium and therefore the sequence is La→Gd. The rare earths are divided into 2 groups.

La - Nd
Sm - Gd

In the first group, Sm impurity is <0.3%. In the 2nd group the Nd impurity is <0.3%. During the La-Ce separation, the Ce content is <0.1%. They can obtain 99.99% La. During Nd-Pr purification they can obtain >99% Nd. During the Sm-Gd separation, the purity of samarium oxide is >99% and the purity of Gd separated is also >99%.

New reagents are being developed and tested for improved rare earth separation and mathematical modelling of the process is of concern.

There is also a development group responsible for the building of equipment. They have worked on lab scale pulse columns and they have examined the possibility of using centrifugal contactors. One major function is that of transfer of technology to the rare earth plant at Baotou.

Pyrometallurgy

They do much work at the Institute on purification of oxides in the pyrometallurgy area and are well equipped to produce many pure products. They are using vacuum arc (La Ce Pr Dy); vacuum melting (Sm and others). They have a high inductance furnace, fused salt electrolysis using a graphide anode. Mixed fluoride systems of La Ce Pr Nd are treated in a $LiF + BaF_2$ bath and produce a purity of 99.9%.

A certain amount of their work is involved with physical metallurgy and applications towards magnets and steels.

Rare Earth Metals Plant, Baotou, Inner Mongolia

This is a 5000 tpy operation treating a slag concentrate, resulting from an iron ore plant about a mile away. The material is transported by pipeline from the iron ore plant to two 100-ft diameter thickners. The underflow grades about 15% rare earths, and is sent to the processing plant where it is processed on a set of tables for further concentration to 30%. From there the material is floated in 22 rougher and 40 cleaner cells up to 55-60% Re_2O_3. They have two types of feeds - monazite and bastnasite.

The products consist of metal chlorides and individual metals and rare earths. The rare earth chlorides account for 3000 t/y. and amount to 500 compounds. The plant is the largest such plant in China, and may be the largest in the world. They have both acid and alkaline processes. The acid process is a pug leach in H_2SO_4, while the alkaline process is a decomposition of the ore at 140°C followed by HCl leach. From the chloride route they produce individual rare earth metals and compounds. Solvent extraction is the method of purification, with products ranging from 99 to 99.9% pure. They plan to produce more compounds in the future, at high purity, by addition of further processing steps. They have used DEHPA in the past, but there is a strong possibility they are using the phosphonic extractant now.

They export 1000 tons of rare earth chlorides (combined and individual) annually. Economically they said that the sulphate route is possibly slightly less expensive than the chloride route. The markets were claimed to be good, but at the moment the rare earth mixtures were more in demand. They have the light fraction as well as the intermediate rare earths.

The 60% concentrate from the flotation circuit goes into the alkaline leach using 40% NaOH, at 30% solids, for a period of 3 hours at 140°C. Unfortunately I was only able to see the concentration and leaching plants as the remainder of the plant, including solvent extraction, was down for maintenance. (The plant was quite old, having operated for about 20 years).

YaoLung Rare Earth Chemical Plant (near Shanghai)

The chemical plant processes monazite to extract rare earths, as well as thorium and uranium. Originally it was a small chemical plant of 23 personnel in 1949 doing some simple chemical conversions. In 1953 they started processing cobalt, and set up a small section on the Cl_2-oxidation method. In 1957, some work was started on monazite for the extraction of thorium and then uranium. This was by classical precipitation methods. Work on rare earths followed.

Sales were difficult at that time, so production was concentrated on mischmetal for lighter flints. The production route was by fusion and electrolysis. They had a gradual expansion of their operations, and the lab work on rare earths started by using IX for isolation. They realized the future importance of rare earths, and the decision in 1968 was made to build a new plant. At this plant they have considered all effluent discharges. They started some work for isolation of Ra^{226}, but found no market. At the moment the tailings discharge is being treated by neutralization. Recently a new yttrium plant was built and they are in the process of moving from the old facility.

The monazite feed to the circuit contains 0.2-0.3% U, 6% Th and rare earths (no values available). The treatment involves the alkaline roast breakdown method, a preferential leach where the Th is dissolved in HNO_3 and the U and Th extracted by TBP. Phosphate is a by-product in this flowsheet. Rare earths separation was done by IX formerly, but now it is SX, using mixer settlers.

The production activities are divided into 5 sections:

1. Ore Dressing

The object is to upgrade the monazite for further purification. Magnetic separation is being used. Recovery of zircon and part of the ilmenite purities is done prior to the NaOH attack.

2. Ore Breakdown

The classical NaOH treatment and chemical separation schemes are being used, where $Th(NO_3)_4$ is being produced and the uranium product being $(NH_4)_2U_2O_7$. This leaves the rare earth nitrate for further refinement.

3. Conversion of Rare Earths

This is the conversion of rare earth chlorides to crystals. There are many separation stages required to produce pure La Ce Pr oxides. Not all rare earths are isolated in this plant, as some are sent to another plant (owned by the company) for special separations.

They export greater than 100 tonnes/year of lighter flints from this plant, also Y and Eu oxides for TV phosphors.

4. Production of Rare Earth Fluorides from Rare Earth Chlorides

5. High Purity Products

This is on small scale under control of the research department for applications research, mainly on phosphors. The work covers the flowsheet aspects for improvement as well as application. They use small batteries of mixer-settlers for their investigation as well as centrifugal contactors.

They have good analytical and instrumentation for quality control and evaluation of products.

6. At their other plant, in Shanghai, they are producing metal (flints), Co-Sm magnets, also Pr-Nd separation and produce separate Pr and Nd oxides and metals.

Purity

Regarding purity, the oxides are for phosphors, and they are required to be very pure. The Y and Eu, which are being marketed in Japan and other countries must be >99.99% pure. Pr and Nd are used for other applications, such as in glass, and the high quality is not demanded, being >95% Pr and 75-80% Nd, depending upon requirements. For samarium oxide, mainly for magnets, not greater than 95% required, but they have produced up to 99.9% purity. They also sell Sm outside China.

As for pigments, Pr mixed with zircon produces a yellow color used for glazing.

The plant has 1500 personnel, of which 100 are comprised of engineers and chemists. They work on a 6-day week. They did at one time have columns, which have since been replaced with mixer settlers. These columns were 6 meters high and 400 mm diameter. At one time they used columns for uranium and thorium, but now used mixer settlers.

They were very interested in crud and flocculents, and particularly interested in solvent-in-pulp and resin-in-pulp, which they would like to be able to apply to their circuit.

In the tour of their pilot plant, they have 2 large banks each of 80 mixer settlers. Each mixer has a capacity of about 2 liters. Into these they fed in about 500 mL feed, using 5 minute retention time. They are using DEHPA, and the feed tanks were on the floor above.

The new Y plant had 2 circuits, each divided into 4 banks of 20 each, or a total of 160 cells. An additional 160 cells were in another area of the building. One motor, with series of belts to mixers, drove all the mixers of each group of 20 cells. The reason for the five minute time in mixing was to increase the SF in the DEHPA-HNO_3 system.

In the analytical lab, they were well up on control, using a UV spectrophotometer for the examination of the rare earths, particularly Y and Yb. They had a Hitachi HPF ultraviolet unit. They had 3 emission specs with ICP attachment, several AA units, particle size analyses (Japanese) and XRF.

The plant had an excellent display room depicting all products by the company. It was quite obvious that the company is capable of producing any type and purity of rare earth compound which the market demands.

University of Fudan, Shanghai and Institute of Organic Chemistry, Shanghai

Visits to both these institutions indicated that considerable rare earth research was in progress, particularly in the development of better organic selective extractants, and in the applications of rare earths to metallurgical and other uses. At the university, research has been connected with the development of fluorescent uses for rare earth mixtures. This research has resulted in the transfer of that technology to industry, and a plant is now producing fluorescent lights incorporating rare earths. The new lights are much closer to daylight intensity than the conventional fluorescent lights.

DISCUSSION OF PAPER BY G. RITCEY

No Notes Taken.

EVALUATION OF TAILINGS STABILIZATION METHODS

By

C.M. Ritcey, and J.M. Skeaff
CANMET
Ottawa, Ontario

And

V.I. Lakshmanan, and L.M. Luckevich
Ontario Research Foundation
Mississauga, Ontario

Canadian Uranium Producers' Metallurgical Committee

1982 Meeting (May 19, 20)
Bancroft, Ontario

ABSTRACT

Four commercial waste solidification processes (based on ironganic and asphaltic binders) were applied to uranium (and leach) mill tailings. The solidified tailings were evaluated by measuring their physical and chemical properties immediately after solidification and as a function of time. The method based on an asphaltic binder was superior but was more expensive by an order of magnitude than a lime based method. Hence the lime based method was considered to be the best compromise, yielding relatively good properties at reasonable costs.

INTRODUCTION

Many types of metal extraction processes, including uranium extraction, produce tailings that require disposal. In general, mine-mill tailings are disposed of by:

- impoundment on land
- dumping into marine or inland waters
- use as a construction material
- use as a mine back fill

However, uranium milling is unique in that it requires the handling and disposal of large quantities of low-level radioactive materials. The special hazards presented by uranium tailings such as radioactivity, acidity and toxicity may prevent direct disposal by these usual methods. Proper management of uranium tailings involves environmental control with respect to:

- Ground water seepage
- Surface water run-off
- ^{222}Rn Diffusion
- Air-borne dust (dry), and
- Dam breakage

The Ontario Research Foundation, in a CANMET sponsored project, (DSS contract serial No. 0SQ80-00131) has assessed alternative potential methods of tailings management as applied to uranium milling. More specifically, the focus of the program was to investigate the stabilization (via solidification) of uranium mine-mill tailings so that impoundment on land is environmentally acceptable. The objectives of this program were to:

1) Acquire information and review waste stabilization technology as applied to uranium mine tailings
2) Acquire and characterize uranium mine tailings
3) Solidify uranium mine tailings by applying existing solidification technology
4) Evaluate the solidified tailings with an assessment emphasis on radionuclide release

2. BACKGROUND INFORMATION

2.1 Origins and Nature of Uranium Tailings

Uranium is extracted from ore by either acid (H_2SO_4) or alkaline (CO_3^{2-}) leaching. After extraction, residual uranium, pyrite, metals (Se, Pb, As, Cd, Mo), inorganic salts (F^-,NO_3^{2-}, Cl^-), and several of decay products are found in the leach residues. The radioisotopes of most concern are ^{226}Ra, ^{230}Th and ^{210}Pb. Systamatic studies by various workers has shown that the majority (80-90%) of the ^{226}Ra is in the fine fraction (-200 mesh) of the tailings solids. Pyrite and pyrrhotite cause problems in the long-term because their oxidation products (H_2SO_4 and Fe^{3+}) solubilize radionuclides and heavy metals, hence increasing their migration to the surrounding environment.

Short term management of uranium tailings usually consists of impoundment behind dams or embankments. Alternatives to impoundment (marine/fresh water disposal, use as construction material or as mine backfill) are not yet available because of their radioactive nature.

2.2 Waste Stabilization Technology

Waste stabilization denotes the conversion of toxic components in a waste to a chemical form that is resistant to leaching. Waste solidification denotes rendering of a liquid or semi-solid waste into a cohesive solid. Often waste solidification results in waste stabilization and vice-versa. In the case of uranium mine tailings, both stabilization and solidification are necessary.

Waste solidification methods applicable to low or medium toxicity wastes can be grouped into the following categories:

- silicate and cement based (inorganic binder)
- lime based (inorganic binder)
- thermoplastic based (asphalt)
- thermoset polymer based
- encapsulation techniques

Organic binder

The advantages and disadvantages of each of these processes is outlined in a table published by Poja Sek.[2], presented here as Table I. Beacuse of the large volumes of tailings requiring disposal and because they are of low radioactivity only the techniques based on inorganic ans asphaltic binders were originally considered as economically feasible for the present program.

Information regarding available waste solidification processes based on inorganic and asphaltic binders was obtained. Four of these methods were selected to be applied to Denison mine tailings; two were lime based, one was silicate based and one was thermoplastic based. The methods used were:

- AECL Bituminization (thermoplastic based) CANADIAN
- Canadian Waste technology (lime based) CANADIAN
- Chemfix (silicate based) U.S.A.
- Pos-o-Tec (IUCS) (lime Based) U.S.A.

A brief discussion of each of the above is outlined below.

2.2.1 Chemfix

The chemfix process is a patented[3] system based on reactions between soluble silicates and silicate setting agents. The mobile reaction unit employed is capable of treating 100,000 gallons or more of sludge per day at ambient temperature and pressure. The process consists of three stages:

i) waste homogenization
ii) Pumping of the waste to the reactor, mixing of waste and reagents, and discharge to disposal area,
ii Solidification within 72 hours

The solidified product has soil-like properties and can be handled by standard earth moving equipment.

2.2.2 Canadian Waste Technology (CWT)

The Canadian Waste Technology (CWT) solidification process is a patented system which "activates" silica in waste sludges so that a complex metal silicate matrix is formed. The silica is present as clay minerals and sand. Treatment of such sludges with a least one of H_2SO_4, HCl, HNO_3, CaO, MgO, $CaCl_2$, $Ca(OH)_2$ and $Mg(OH)_2$ (as determined by Canadian Waste Technology) yields a pourable slurry which solidifies over an unspecified time period.

2.2.3 Pos-o-Tec (IUCS) Process

The I.U. conversions Systems (IUCS) Pos-o-Tec process is a patented[4] micro-encapsulation process that involves the precise mixing of waste products, fly ash and lime-rich materials. The treated waste has the consistancy of dry soil-cement and is compacted using vibratory rollers. After curing, compressive strengths of 1000 psi and permeabilities of 5×10^{-9} in s^{-1} have been recorded. It should be noted that the treated waste is not pumpable.

2.2.4 AECL Bitúminization

Bitúminization consists of mixing the waste with asphalt, (a petroleum byproduct composed of high molecular weight hydrocarbons), at 165 - 167°C. Water originally present in the waste is volatilized at this temperature. In this particular case, the tailings were dried and ground prior to bitúminization.

3. DESCRIPITION OF TAILINGS AND SOLIDIFICATION METHODS TESTED

3.1 Tailings Characterization

Two barrels of mine tailings were received from Denison Mines located at Elliot Lake, Ontario. In the Denison uranium recovery process, the acid leached slurry is hydrocycloned and overflow is discharged to thickeners. The thickener overflow is discharged to an unclarified pregnant storage tank; the thickener underflow is combined with the hydrocyclone underflow and filtered. The filter cake is washed in two stages of drum filters. The second stage cake is washed with water, reslurried with neutralized barren solution and pumped to the tailings disposal area. The tailings are aprroximately 80% solid.

A pipe sample of the tailings was submitted for sieve, X-Ray fluorescence (XRF) and X-ray diffraction (XRD) analsis. Based on the ^{226}Ra analysis (352 pCi/g) and the sieve analysis (Table II), the sample of Denison tailings received can be considered typical. Hence, it is presumed that elutriation of fines containing major portions of the ^{226}Ra did not occur. Analyses by XRF and XRD are given in Table III and IV respectively. Since the pyrite analysis by XRD and XRF were relatively low, the tailings were analysed using atomic absorption spectroscopy. The results indicated 2.06 percent iron, thùs giving an upper limit on the pyrite of 4.4 percent. These values correspond to those expected for Denison talings.

4. EXPERIMENTAL MIXES

In all of the solidifacation experiments the actual tailings solidification was performed by process vendor personnel. Three of the experiments were performed at O.R.F. with O.R.F. personnel observing. The AECL Bituminization was performed at the AECL Chalk River Laboratories. No O.R.F. personnel were present but a detailed description of the solidification was provided by AECL.

4.1 Chemfix

Two samples of uranium mine tailings were solidified using the Chemfix process. The first, designated C1, mixed 18.0 kg of tailings (84.5% solids) 1.7 L H_2O, 0.74 kg. inorganic setting agent and 0.18 L type N sodium silicate solution. The second, designated C2 contains 12.0 kg of tailings (84.5% solids), 0.74 kg inorganic setting agent, and 0.60 L H_2O. After mixing 6.0 kg more of tailings were added. C1 was a thixotropic slurry, easily placed by hand; C2 was a stiff slurry, readily placed using a small concrete vibrator.

4.2 Canadian Waste Technology (CWT)

Two samples of uranium mine tailings were solidified using the CWT process. The first, designated KC (Krofchak cast), mixed 18.0 kg. tailings (80% solids), 2.6 kg water and 0.9 kg reagent for 5 minutes in a standard concrete mixer. The second, designated KF, (Krofchak Filtered), had an increased water content (to give a 30% solids slurry) and was filtered through a Perin plate and frame filter press. Mix KC was a thixotropic, pourable slurry easily placed by hand; mix KF was a thixotropic, stiff slurry readily placed with mild vibration.

4.3 I.U. Conversion System (IUCS) Process

Two samples of uranium mine tailings were solidified using the IUCS process. The first, designated IU1, mixed 1L H_2O, 4.8 kg of IUCS blend No. 1 and 4.267 kg. tailings (84.5% solids). The second, designated IU2, mixed 0.6 L H_2O, 3.524 kg IUCS Blend No. 2 and 6.101 kg tailings (84.5% solids). Both IU1 and IU2 were extremely dry and required vigorous hand compaction to form a continuous body; the moisture content of the solidified tailings was estimated to be 15-16%.

4.4 AECL Bituminization

The uranium mine tailings were dried (110°C) overnight and ground. The dry powder was screw fed (10-11 kg/h) to a bituminizer (ASTM type # Roofing Asphalt) with a bitumen flow rate of 3 to 3.5 kg/hr. Production rate ranged from 13 to 14 kg/hr with the final product being 75 to 76 weight percent tailings solids. The product was discharged to 11 L wooden moulds; each sample weighed 20-21 kg.

5. EVALUATION OF SOLIDIFIED TAILINGS

5.1 Physical and Mechanical Evaluation

5.1.1 Flow Properties of Material for Placement

The consistency of the materials will determine the type of placement method and to a large extent, the product strength and durability. The consistency of cement mortars is most commonly measured by Flow Value (ASTM C109-77). This method was used for the Chemfix and Krofchak processes. The IUCS material was too dry for the test to be applicable. The flow of the AECL material would, of course, be temperature dependent and no tests were run on this material. The water solid ratios, percentage dry tailings and flow properties of all mixes are presented in Table I.

5.1.2 Shrinkage, Compressive Strength and Flexural Strength

The shrinkage of stabilized tailings will be important in determining the extent of break-up of a large stabilized mass of tailings. The initial movement as the samples dry or the stabilizing reactions take place, will be the most important in this regard. Prior to demoulding the samples, measurements were made of the shrinkage which the samples had undergone and this is expressed as a percentage of the original length in Table VI.

The measurement accuracy was 0.1 percent, that is the AECL sample showed no measureable shrinkage at the time of demoulding. The IUCS samples had been cured, covered with a plastic sheet until demoulding while the CWT and Chemfix samples had been covered for the first week of curing and thereafter left exposed in normal laboratory air. All of these curing conditions were arrived at in consultation with the representatives of the companies involved in the stabilization. Prior to demoulding, all Krofchak Cast (KC) samples developed large open shrinkage cracks. This was not observed in the Krofchak Filtered (KF) samples although they also shrank considerably, see figure no

Compressive and flexural strength were measured at 14 or 28 days and are presented in Table VI as well. Flexural testing was performed on 3" cubes as described in ASTM C 293-77 "Flexural Strength of Concrete (Using Simple Beam with Center-Point Loading)". Samples were cut from the cast specimens with a nominal 3" x 3" cross-section and were tested over a 12" span. The compressive tests were performed as described in ASTM C116-68 "Compressive Strenght of Concrete Using Portion of Beams Broken in Flexure". Samples were nominally 3" cubes. The AECL samples did not have a well defined failure point, but were loaded to 10 percent deformation.

5.1.3 Freeze-Thaw Resistance

All samples were subjected to freeze-thaw cycling by ASTM C666-77 "Standard Method of Test for Resistance of Concrete to Rapid Freezing and Thawing". Procedure A which involves freezing and thawing the samples immersed in water was used for all tests. Samples were curred for 14 days (AECL and Chemfix) or 28 days (IUCS or CWT) prior to testing. All samples were immersed for a minimum of 24 hours prior to beginning freeze-thaw cycling. An attempt was made to determine the length change of the specimens on freeze-thaw cycling, but this was not possible due to surface scaling, or sample disintegration or warping. The results are summarized in Table VII. No quantitative comparison of the freeze thaw samples was made. A qualitative classification of the freeze-thaw resistance (from observations) is that AECL is better than IUCS which is better than Chemfix which in turn is better than CWT.

5.1.4 Porosity and Pore Size Distribution

Measurements of porosity and pore size distribution were performed on the seven solidified tailings samples using an Aminco 60,000 psi mercury intrusion porosimeter. Intrusion volumes were measured as a funciton of of both increasing and decreasing pressure. The shapes of the hystereses curve can be related to the shape of the pores found in the solidified matrix The total pore surface area of each sample was calculated using the acquired data. Porosity data are given in Table VIII.

The data indicated that the porosity of the samples varied over the range 4.9 to 39.7 volume percent. Reduction of the water used in preparation of the Krofchak and Chemfix samples gave the expected decrease in total porosity. The AECL bituminization process resulted in a very low porosity material. The IUCS material was found to have pores in two ranges, the smaller sized pores presumabley associated with the added pozzolan. When this is taken into account, the pore size distributions of the six materials solidified with inorganic agents were quite similar, except for the expected shift to samller pore sizes observed with decreasing porosity. The materials solidified by IUCS and Chemfix both appeared to have a higher fraction of partially blocked pores in the gel structrue than the Krofchak sample as illustrated by the more pronounced hysteresis effect observed in their pore size distribution curves.

5.2 Radon Emanation

One of the important parameters in the evaluation of the effectiveness of fixating processes for uranium tailings should be the rate of radon gas emanation from the fixed materials. For a block of fixed material which contains a percentage of uranium mine-mill tailings, the radon will emanate from the surface layer and from the interior of the block. The depth within the block from which the gas originates is dependent upon the diffusion constant of the gas through the material.

A procedure was designed to estimate the rates of radon emanation from the seven fixed materials and also from untreated tailings. The fixed materials were prepared in the shape of cubes having a volume of 0.44 litre. Each of the samples was placed in a 4 litre air-tight steel box fitted with a sealing connector and a one-way valve. To obtain each radon gas sample an evacuated radon cell having a volume of 160 mL was connected to the sample chamber, thus transferring a sample of gas to the cell through a 0.8 µm pore size Millipore filter. This procedure filtered out the airborne radon daughter products from the gas sample. The gas cell was disconnected, and after a 2-3 hour period, a ten-minute count was made of the activity in the cell. Samples of gas were taken and counted at regular daily intervals over a period of a week, in order to determine the trend of the radon gas concentration as a function of time.

The data was corrected for background counts and for the dilutions during the sampling. A graphical representation of these data showed that parallel relationships of radon concentration versus time were obtained. The counts that were acquired on the last day of each run could therefore be used to generate a comparison between the fixed materials. Table IX shows the results of the radon emanation measurements. The two right-hand columns represent the radon counts per unit weight of fixed sample and per unit weight of tails respectively, with both sets of data normalized to a reading of 100 for the untreated tailings.

From this simple test procedure, the AECL bituminization was most effective in reduction of the rate of radon gas release, while the Krofchak poured sample (KC) allowed the greatest release of radon gas. The two IUCS samples were next best at hindering radon release. The filtered sample (KF) and the two Chemfix procedures of fixing were less effective, and gave approximately the same results. However, when compared on a per unit weight of tails basis, only the AECL bituminization actually reduced the radon release relative to that of the untreated tails. When comparing on the basis of a unit weight of fixed material, both the AECL and IUCS reduced the radon gas emanation below that of the untreated tails. Because of the variability of the water content among the samples, the aging of the samples over the two week measuring period, uncertainties in the normalization and statistical uncertainty, the total uncertainty associated with each of the relative radon counts per unit weight is estimated to be about 20%.

It seems anamalous that the radon counts per unit weight of tailings are higher than the count for the untreated tailings for all of the samples but one. The histories of the seven fixed materials prior to the radon counting are similar in that a period of two to four weeks elapsed between formation and the time at which the measurements were taken. During this period, the fixed materials aged at their own specific rates. The aging of the untreated tailings occurred over a much longer time. The difference in radon release rates between the group of seven fixed samples and the untreated tailings due to the different time periods involved is not known.

Also, because radon dissolves readily in water, the percentage of water in the seven fixed samples and the untreated tailings could result in a significant variation in the radon diffusion constants of the samples. Since these factors are not known, a direct comparison of the radon data between the group of seven fixated materials and the untreated tailings may not be possible.

5.3 Chemical Durability

Tests carried out to study the chemical stability of the various fixated materials (after crushing through jaw crusher to produce $^3/4$ inch materials) include agitation and percolation leach studies for the release of ^{226}Ra.

5.3.1. Agitation Leach Studies

Agitation leach tests with various fixated materials were carried out in 1 L leach vessels at 10% solids, one-hour residence time, room temperature and pH levels of 1.5, 3.0 and 5.5. Sulphuric acid was used to adjust the pH of the leach solution; agitation rate was 98 rpm.

The slurry samples at the end of the test work were filtered to obtain samples for ^{226}Ra analysis. The solution samples were stored in containers previously washed with distilled water and nitric acid. In order to prevent loss of ^{226}Ra values from solution, the samples were stabilized by the addition of 10 mL conc. HNO_3/L filtered solutions. Leach residues were subjected to screen analysis. ^{226}Ra in solution was obtained by an α-spectroscopic method.

A large number of agitation leach studies investigation the effects of particle size, pH and fixation method on ^{226}Ra release were performed. A summary of data only will be presented and is given in Table X.

The lowest solution ^{226}Ra levels were obtained for the IU2 solidified tailings (< 100 pCi/L). This may be due to the alkaline pozzolan reaction binding ^{226}Ra within the crystalline lattice i.e. with ^{226}Ra substitution for Ca occurring during precipitation of cementitous compounds.

Increasing the surface area of the solidified tailings (i.e. decreasing the average particle size) does not increase ^{226}Ra release. This is possibly due to simultaneous sorption, desorption and precipitation of aqueous ^{226}Ra and depends on complex solution equilibria.

Both decreasing the pH (to 1.5 from 3.0) and increasing the pH (to 5.5 from 3.0) increase ^{226}Ra release. Fluctuations in the pH affect the complex solution equilibria in various ways either precipitating or dissolving radium sulphate. It is however, emphasized that comprehensive testing was not performed and that the data can only be interpreted within the limited scope of the experiments.

5.3.2 Percolation Leach Tests

It is extremely difficult to simulate the seepage of rain water through natural and various fixated tailings piles on a laboratory scale. Nevertheless, in order to obtain an estimate of the total radium that could be released through seepage, a number of one inch diameter containers were packed with various fixated materials and tailings. The rate of flow of seepage was simulated to correlate with total precipitation in the Eliott Lake area (34''/year) and the pH of the water was maintained at 4. Based upon the pH of rain water east of the Mississipp River in North America (4) the pH of the leachant was chosen at 4.

Seepage samples were collected daily and the pH of these seepage samples was measured and the samples were then preserved by the addition of 10 mL conc. HNO$_3$/L. ^{226}Ra levels in these solutions were measured by α-spectroscopic methods.

The test results (Table XI) generally show a decrease in ^{226}Ra with an increase in time. Based upon the seepage analysis, effluent samples obtained from IU2 fixated material contained the lowest total ^{226}Ra in solution. The pH of the seepage samples collected for the Krofchak samples were highest at 12.1-12.4 and for the

AECL samples were lowest at 7.0-8.2. This should be compared against the pH of seepage samples (2.8-3.2) obtained from untreated tailings (Table XII). Percolation leach tests during the test period with blank tailings were not successful. Sufficient daily seepage samples could not be collected since leachant solution did not seep through.

6. DISCUSSION

As can be seen from the results given above, considerable variation is observed in the performance of the four solidification processes. The following is a summary of some of the advantages and disadvantages of these processes, including some preliminary cost information.

The solidification process developed by Canadian Waste Technology was the least expensive of the processes listed. No estimate has been made of capital costs, but materials costs would be in the order of $1.50 per ton of solidified tailings. The test results indicate, however, that tailings solidified using this process exhibit poor strength, shrinkage and freeze-thaw resistance. As would be expected, this results in relatively high rates of radon emanation and acidic leach rates of radium, when compared to other processes tested. It would appear that the process suggested by Canadian Waste Technology may not be suitable for the solidification of tested uranium mine-mill tailings.

The bituminization process performed by AECL resulted in solidified tailings with very good physical properties and very low radon emanation rates. The product was found to have relatively high radium leach rates, this being somewhat surprising in light of the good physical properties. The results of the leach studies are quite significant since control of the radium and radon emanation is the primary purpose of the fixation procedure. Since in the Canadian climate precipitation rates exceed those of evaproation, the constant leaching of ^{226}Ra is of special significance. As a result, the bituminization process would be more suitable in a location with a more arid climate since in this case release of radon and surface erosion would be of major concern.

The cost of the bituminization process is its most serious disadvantage. As tested, materials costs alone were approximately $ 420 per ton of solidified tailings. If the bitumen content were dropped to 10% this value would drop to approximately $140 per ton. For cost reasons alone, this type of process is therefore not practical for solidification of the bulk of the tailings. Perhaps it can find some limited application as a solidification process to be used to place a cover over an existing tailings, this cover representing only 1-2% of the total volume of the tailings.

The cost of the Chemfix process has been estimated at approximately 300 to 500 thousand dollars in set-up and $7 per ton of tailings for materials. This process is the second least expensive of those tested. Solidification of tailings using this process resulted in some improvement in strength and porosity values over unsolidified tailings, but resulted in a product with poor freeze-thaw resistance. A distinct advantage over the other processes is its rapid setting rate. Radon and radium release rates were generally in agreement with the physical properties, the values being lower than those observed for the CWT samples but not as low as observed for the AECL samples (except for ^{226}Ra agitation leach rates) and IUCS samples.

The samples solidified using the IUCS process exhibited good strength and porosity values and fairly good resistance to freeze-thaw action. The relatively good physical properties were followed by similar results for the radon emanation and radium leaching studies. The IUCS solidified materials showed the best resistance to acid leaching and second lowest radon release rates when compared to other processes tested. At a cost (materials and processing) of $10-$17 per ton of solidified tailings, this process is considerably less expensive than the bituminization process, but the most costly of the inorganic processes. The IUCS costs assume pozzolan and additive sources relatively close to the site and an existing plant. Another advantage of the process is that placement of the material is performed using conventional construction equipment.

7. CONCLUSIONS

The solidification of uranium mine-mill tailings to minimize risks of environmental contamination can be performed by several different processes. Although the present study is not comprehensive, it does indicate what properties may be expected from tailings solidified using processes of various types. As can be seen by the summary given below, the process marketed by Canadian Waste Technology was the poorest performer of the group tested.

SUMMARY OF RESULTS

	CWT	AECL	CHEMFIX	IUCS
Percent Tailings (approx)	94	75	95	50
Ease of Placement	1	4	2	3
Shrinkage	4	1	3	2
Compressive Strength	4	3	2	1
Flexural Strength	4	1	3	2
Freeze-Thaw Resistance	4	1	3	2
Porosity	2	1	2	2
Pore Shape	3	-	2	1
Batch Leach	4	2	2	1
Percolation Leach	4	2	2	1
Radon Emanation	4	1	3	2
Cost - Materials	1	4	3	2
- Process	1	4	1	3

* Processes are ranked 1-4, 1 being the best performer of the group and 4 being the worst performer.

Processes 1-3 gave the most improved properties of the uranium tailings, but the bituminization process is much more expensive. The silicate based process (Chemfix) showed some improvement but did not perform as well as the lime-pozzolan or bituminization processes. These categories of solidification processes were suggested as being feasible in section

3 of this report. The specific solidification process, however, is not necessarily representative of all processes encompassed by each category.

It is obvious from the results that total solidification of the tailings is not possible, especially considering the costs involved. The solidified materials, however, may be quite useful for containment of the tailings either by capping or erection of barriers.

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2. U.S. Patents 3,841,102 and 3, 837,872

3. U.S. Patent 4,229,295

4. U.S. Patent RE - 29783

TABLE I

ADVANTAGES AND DISADVANTAGES OF SOLIDIFICATION PROCESSES

Process	Advantages	Disadvantages
Cement-based	1. Additives are available at a reasonable price. 2. Cement mixing and handling techniques are well developed. 3. Processing equipment is readily available. 4. Processing is reasonably tolerant of chemical variations in sludges. 5. The strength and permeability of the end-product can be varied by controlling the amount of cement added.	1. Low-strength cement-waste mixtures are often vulnerable to acidic leaching solutions. Extreme conditions can result in decomposition of the fixed material and accelerated leaching of the contaminants. 2. Pretreatment, more-expensive cement types, or costly additives may be necessary for stabilization of wastes containing impurities that affect the setting and curing of cement. 3. Cement and other additives add considerably to weight and bulk of waste.
Lime-based	1. The additives are generally very inexpensive and widely available. 2. Equipment required for processing is simple to operate and widely available. 3. Chemistry of pozzolanic reactions is well known.	1. Lime and other additives add to weight and bulk of waste. 2. Stabilized sludges are vulnerable to acidic solutions and to curing and setting problems associated with inorganic contaminants in the waste.
Thermoplastic	1. Contaminant migration rates are generally lower than for most other techniques. 2. End-product is fairly resistant to most aqueous solutions. 3. Thermoplastic materials adhere well to incorporated materials.	1. Expensive equipment and skilled labor are generally required. 2. Sludges containing contaminants that volatilize at low temperatures must be processed carefully. 3. Thermoplastic materials are flammable. 4. Wet sludges must be dried before they can be mixed with the thermoplastic material.
Organic polymer	1. Only small quantities of additives are usually required to cause the mixture to set. 2. Techniques can be applied to either wet or dry sludges. 3. End-product has a low density as compared to other fixation techniques.	1. Contaminants are trapped in only a loose resin-matrix end-product. 2. Catalysts used in the urea-formaldehyde process are strongly acidic. Most metals are extremely soluble at low pH and can escape in water not trapped in the mass during the polymerization process. 3. Some organic polymers are biodegradable. 4. End-product is generally placed in a container before disposal.
Encapsulation	1. Very soluble contaminants are totally isolated from the environment. 2. Usually no secondary container is required, because the coating materials are strong and chemically inert.	1. Materials used are often expensive. 2. Techniques generally require specialized equipment and heat treatment to form the jackets. 3. The sludge has to be dried before the process can be applied. 4. Certain jacket materials are flammable.

TABLE II

SIEVE ANALYSIS OF URANIUM MINE TAILINGS

SIZE FRACTION (mesh)	WT. PERCENT
+48	4.16
-48 + 65	7.84
-65 + 100	9.23
-100 + 150	2.50
-150 + 200	17.28
-200 _ 325	10.34
-325	48.66

TABLE III

SEMI-QUANTITATIVE CHEMICAL ANALYSIS (XRF) OF URANIUM MINE TAILINGS

SAMPLE I

Element	Approximate Weight Percent
Si	High
Al	3
S	2
K	2
Ca	2
Fe	1.5 (3.2*)
Ti	0.3
Na	<0.1
Mg	<0.1
Pb	<0.1
Zn	<0.1
Rb	<0.1

SAMPLE II

Element	Approximate Weight Percent
Si	35.1
Al	3.3
Fe	2.4 (5.2)*
K	1.69
Ca	1.41
Ti	0.22
P	0.05
Mg	0.07
Zn	Trace
Se	Trace
Cu	Trace
Pb	Trace
Ni	Trace

* % Pyrite if all Fe is present as FeS_2.

TABLE IV

XRD Analysis of Uranium Mine Tailings

Quartz	70 - 80%
Muscovite (or similar Mineral)	10 - 20%
Gypsum	5 - 10%
Pyrite	1 - 2% *
Others present at	<1 - 2%

* Wet chemical analysis yielded a value of 4.4% pyrite.

TABLE V

WATER/SOLIDS RATIO, PERCENTAGE DRY TAILINGS AND FLOW PROPERTIES OF MIXES

	Percent Solids	$\frac{\text{Dry Tailings}}{\text{Total Dry Weight}}$ (%)	Flow Value *	Comments
Krofchak Cast	72.1	94.1	150	Flows readily
Krofchak Filtered	-	94.1	128	Placeable by hand
AECL	75.7	75.7	-	Flows readily at 165°C
Chemfix I	78.3	94.5	136	Easily placeable; almost flows
IUCS 1	83.5	42.9	-	Requires hand or vibratory compaction
IUCS 2	84.7	60.0	-	Requires hand or vibratory Compaction.

* As per ASTM C 109-77.

TABLE VI

SHRINKAGE, COMPRESSIVE STRENGTH AND FLEXURAL STRENGTH OF SAMPLES

Sample No.	Process Used	Shrinkage at Demoulding (%)	Compressive Strength Kpa	Compressive Strength psi	Flexural Strength Kpa	Flexural Strength psi
KC	Krofchak Cast(1)	1.6	435	65	190	30
KF	Krofchak Filtered(1)	1.9	580	85	135	20
AECL	Bituminization(2)	< 0.1	1080	155	1480	215
CI 1	Chemfix 1(2)	0.4	1230	180	425	60
CI 2	Chemfix 2(2)	0.3	2590	375	415	60
IU 1	IUCS 1(1)	0.1	9215	1335	1370	200
IU 2	IUCS 2(1)	0.1	8030	1165	1260	180

(1) All tests 28 days after casting.

(2) All tests 14 days after casting.

TABLE VII

SUMMARY OF FREEZE-THAW TESTS (2 SAMPLES OF EACH MATERIAL TESTED)

SAMPLE		CYCLES AT END OF TESTING	COMMENTS
Krofchak Cast	A	0	Turned to mush during saturation period prior to freeze-thaw cycling.
	B	6	Broke prior to freeze-thaw test; turned to mush after 6 cycles.
Krofchak Filtered	A	6	Broke after 1 cycle; turned to mush after 6 cycles.
AECL	A	128	No visual change of surface; some warpage
	B	128	No visual change of surface; some warpage
CHEMFIX 1	A	11	Ends disintegrating; severe spalling on one side
	B	11	Ends disintegrating; one side turning to mush.
CHEMFIX 2	A	5	Numerous cracks; shattered
	B	5	Numerous cracks; shattered
IUCS 1	A	40	Severe spalling and disintegration
	B	80	Severe spalling and disintegration
IUCS 2	A	98	Severe spalling and disintegration
	B	98	Severe spalling and disintegration

TABLE VIII

SUMMARY OF PORE SIZE ANALYSIS DATA

Sample No.	Process Used	Density at 1 psi (kg/m^3)	Density at 60,000 psi*	Measured Porosity (Volume %)	Absolute Intrusion (cc/g)	Major Pore Size Range	Surface Area (m^2/g)
**	Unsolidified Tailings	1,414	2,369	40.3	0.2850	0.9 - 10.0	0.76
KC	Krofchak - poured Sample	1,611	2,668	39.7	0.2461	0.6 - 3.0	2.45
KF	Krofchak - prepared Using filter press	1,564	2,322	32.7	0.2089	0.4 - 3.0	2.26
AECL	Bituminization	1,790	1,883	4.9	0.0276	1.0 - 5.0	4.70
C1	Chemfix - higher Water content	1,520	2,336	34.9	0.2299	0.5 - 2.5	3.22
C2	Chemfix - lower water content	1,705	2,412	29.3	0.1717	0.4 - 1.5	5.99
UI 1	IUCS - additive to tailings ratio = 1:5	1,600	2,493	35.8	0.2239	(1)0.8 - 4.0 (2)0.02- 0.2	20.92
UI 2	IUCS - additive to tailings ratio = 0.75	1,656	2,529	34.5	0.2085	(1)0.4 - 3.0 (2)0.03- 0.2	16.44

* Density of tailings alone based on composition given in Table 4 - III is approximately 2.70 g/cc.

** Measurements performed on dried tailings, no crushing or ginding being performed on the dry product.

TABLE IX

RADON EMANATION DATA

SAMPLE	MASS OF SAMPLE g	WEIGHT DRY TAILS/ TOTAL DRY WEIGHT %	RADON COUNT*	RADON COUNT PER UNIT SAMPLE MASS**	RADON COUNT PER UNIT TAILS MASS**
KC	719	94.1	43346	270	287
KF	679	94.1	27121	179	190
AECL	776	75.7	884	5	7
C1	680	94.5	26230	173	183
C2	740	94.5	33824	205	217
IU1	704	42.9	8299	53	123
IU2	762	60.0	13618	80	134
TAILS	243	100	5420	100	100

* 10 minute count that has been corrected for the background and the dilutions when sampling

** Data has been normalized to 100 for untreated tails

TABLE X

SUMMARY OF AGITATION LEACH DATA

Fixation Method	Initial Ra-226 conc. (pCi/g)	pH = 1.5		pH = 3.0		pH = 5.5	
		-3/4" + 1/2"	-1/4"	-3/4" + 1/2"	-1/4"	-3/4" + 1/2"	-1/4"
		Ra-226 (pCi/L)	Ra-226 (pCi/L)	Ra-226 (pCi/L)	Ra-226 (pCi/L)	Ra-226 (pCi/L)	Ra-226 (pCi/L)
KF	334	545	-	122	208	305	109
KC	355	245	-	90	207	178	145
C1	305	161	210	176	80	97	28
C2	298	218	412	140	158	76	70
IU1	144	103	124	60	28	72	16
IU2	212	85	102	29	55	60	16
AECL	277	195	329	56	39	340	91
Denison Tailings	352	-	360	-	426	-	374

TABLE XI

PERCOLATION LEACH TEST RESULTS

Conditions: Leach Solution pH 4.0
Flow Rate: 0.170L/Day
(Equivalent to 34"/year)

Fixation Method	Head pCi/g	Day 1 ^{226}Ra pCi/L	Day 8 ^{226}Ra pCi/L	Day 15 ^{226}Ra pCi/L	Day 22 ^{226}Ra pCi/L
Krofchek, Filtered	334	823	474	462	488
Krofchek, Cast	355	668	641	488	383
Chemfix 1	305	271	284	220	210
Chemfix 2	298	256	238	150	
IU 1	144	70	31	53	69
IU 2	212	47	17	27	9
A.E.C.L.	277	62	80	74	100

TABLE XII

Energy, Mines and Resources
Final Report

TABLE 5-XXVI

Effect of Time on the pH of Various Percolation Seepage Samples (1)

	KF	KC	C1	C2	U1	U2	AECL	UT(2)
Day 4	12.1	12.2	9.0	10.7	11.0	9.3	7.0	2.9
Day 7	12.1	12.2	10.6	11.2	11.0	10.0	7.9	3.2
Day 10	12.3	12.3	10.6	11.3	11.0	10.0	7.8	2.8
Day 14	12.4	12.4	10.2	11.0	11.0	9.9	7.6	3.0
Day 17	12.3	12.3	10.4	11.0	10.9	9.7	7.0	2.9
Day 20	12.4	12.5	10.2	10.9	10.8	9.7	8.2	2.9
Day 24	12.4	12.4	10.4	10.9	10.8	9.7	7.0	2.8
Day 27	12.4	12.4	10.2	10.8	10.6	9.8	7.5	3.0
Day 30	12.2	12.2	10.3	10.6	10.4	9.6	7.7	—

(1) Initial pH of leachant solution, 4.0.

(2) Untreated tailings.

DISCUSSION OF PAPER BY V. LAKSHMANAN

P. Fournier	Was the drying cost included in the costs?
V. Lakshmanan	Yes it was. The drying temperature was 110°C.
I. Itzkovitch	What acid was used to acidify the leach tests at pH 1.5?
V. Lakshmanan	H_2SO_4.
Question	What is the estimated cost of the process?
V. Lakshmanan	Chemfix $7/ton Bitumen $420/ton Krafchak $1.50-2.00/ton IUCS $10-17/ton
L. Heymann	If you had used cement, it should be less than $7.00.
V. Lakshmanan	That is essentially the Chemfix process.
W. Craigen	Are the costs just the chemical costs?
V. Lakshmanan	They are the process costs.
L. Heymann	Is stabilization necessary or is it of intellectual interest?
V. Lakshmanan	No, it may be necessary to cap the tailings, so stabilization methods have to be evaluated.
I. Itzkovitch	I believe that the information is of importance in estimating base line data for viability and cost.
M. Campbell	Last year, at the Fort Collings meeting on Tailings, was discussed the remedial action in the USA for 24 tailings stockpiles. The estimated cost was $380 million. Therefore we must know what are the best possible processes to use. For example, whether tailings were to be physically moved or were to be capped. A congressional subcommittee report was issued by the US government and is available from me for those interested.
V. Lakshmanan	Perhaps Dr. Gee would like to comment on the USA situation.
G. Gee	I am involved with some of the USA programs with DOE. Much work is in progress on technical development but no work is actually started at any of the sites. After examination of the possibility of moving sites, only 1 site of 40 hectares in Salt Lake City is under debate and consideration for moving. But the cost will be $70 million. The actual cover material for sites has not been decided. However, a thin cover of asphalt emulsion, to seal radon, is now being considered.

Energy, Mines and Resources Canada
Énergie, Mines et Ressources Canada

CANMET
Canada Centre for Mineral and Energy Technology
Centre canadien de la technologie des minéraux et de l'énergie

COMPOST PRODUCTION FROM MUNICIPAL WASTES OF CANADIAN MINING TOWNS

A. Jongejan
Extractive Metallurgy Laboratory

June 1981

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COMPOST PRODUCTION FROM MUNICIPAL WASTES OF CANADIAN MINING TOWNS

by

A. Jongejan*

and Presented by G.M. Ritcey**
CANMET, Ottawa

ABSTRACT

A summary of results of experiments on composting municipal wastes, and an overview of a type of composting process that could be used in small Canadian mining towns are given. The process is a means of waste disposal designed to produce compost. Compost can be used for the revegetation of mine-mill tailings as its sorptive properties complement the chemical action of inorganic fertilizers. The possibility of using compost instead of peat in water pollution-abatement processes can be considered.

Difficulties that can be expected if a windrow composting process is continued during the low ambient-temperatures of Canadian winters can be avoided by storing the garbage-sewage mixture as hydraulically-compacted briquettes. Degradation of the briquettes takes place during mild-temperature periods without producing the foul odors of heaped garbage.

A tentative plan for a composting plant is presented as an illustration of the application of the experimental results in a practical process. Because the process is a means of waste disposal, costs have to be divided between the municipality and the mining industry.

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* Research Scientist, and **Head Solution Purification Section, Extractive Metallurgy Laboratory, CANMET, Energy, Mines and Resources Canada, Ottawa, Ontario.

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INTRODUCTION

Pollution of the environment originates to a great extent in the production of wastes that become part of the natural environmental processes. Wastes are not only produced by industry, but also by the human population, its institutions and enterprises generally not considered to be "industrial". Because natural processes in the environment do not discriminate between the sources of waste, the solution to environmental problems should be based on cooperation between different groups of society. Generally, the following can be considered from geochemical points of view. Natural processes are cyclic. The concept of "recycling" materials and energy developed in the last few decades when problems concerning energy shortage, waste disposal and pollution of the environment arose.

Natural resources consist of inorganic and organic materials. As compared with the direct recovery of energy from sunlight, wind, water and earth's heat, energy is indirectly provided by materials of the ancient and present biosphere. Fossil fuels represent the preservation of solar energy in the earth crust and can be compared with the fuel supplied by the present biosphere in the form of wood from the forests and other materials from plants and animals (life energy in the food chain).

In addition to energy production the following distinctions can be made based on the nature of the materials to be recycled from wastes:

a. "Recycling" of inorganic materials and derivatives of fossil fuels, e.g., ceramics, metals, petroleum and coal products;

b. "Recycling" of organic materials connected with the present biosphere (conversion of products of plant and animal life to compost).

Materials of cycle "a" may be cycled several times and will be returned to the earth crust at a rate which is comparatively slower than those of cycle "b" which coincides with the rate at which they are produced.

The treatment of wastes covers therefore, i) "material" recovery (recycling of inorganic materials), ii) "energy" production (recycling of energy preserved by the earth) and iii) "compost" production (recycling of organic materials of the present biosphere). The preferable type of conversion varies per locality, e.g., energy production and material recovery for large cities, compost production for rural areas. The combination of treatments of liquid (sewage) and solid (refuse) wastes is not only preferable from a theoretical, but also from an economical point of view. Although the use of large areas for lagoons in sewage treatment and that of other areas for refuse disposal by the landfill method may seem economical at present, the use of a plant in which both types of wastes are treated will save much in view of land occupation. According to Nell and Wiechers (1) "The utilization of raw sewage sludge during high temperature composting eliminates the need for costly anaerobic sludge treatment". It is to be understood that the raw sludge be dewatered to approximately 60% before composting.

One type of waste produced by the mining industry is the tailings which contains waste sulphides such as pyrite. Rain water and oxidation of this mineral cause acid-water pollution. Because tailings consist of finely-ground waste-rock, the disposal areas are devoid of vegetation. Apart from aesthetic considerations, the tailings dust being blown about by the wind, creates air pollution. In order to abate the air pollution by the tailings dust and for aesthetic reasons, revegetation of the tailings disposal areas is required. This is done by the application of fertilizers and plant seeds. The fertilizers provide the chemical nutrients for the germinating seeds and growing plants. The humus layer required to sustain growth is developed by the slow process of plant decay. The application of compost, therefore, would provide the sorptive medium required by vegetation. The use of compost as a carrier for liquid fertilizers may be considered. "Compost has not replaced chemical fertilizers; instead the two complement each other" (2).

Compost can be produced by various types of organic matter, including municipal wastes in the form of garbage and sewage. Disposal of municipal wastes can be achieved by converting these materials into compost. The situation in Europe is described by Wiley (2): "Indications of the interest in composting are the increasing number of plants; the cooperation between agricultural, public health and public works officials". Although the mining industry produces the pollution, revegetation of tailings-disposal areas required by laws concerning public health can be considered to be an agricultural activity making use of material produced by municipal waste disposal plants usually operated by public works personnel.

Apart from tailings revegetation, the sorptive properties of compost might be considered in connection with water pollution. Compost resembles peat, and processes using peat for the abatement of water pollution have been designed (3,4).

Various processes are used to treat wastes of large urban populations. The principal interest in these cases is in material recovery by recycling (type 'a'; p 1) and in energy production (Refuse Derived Fuel). The organic fraction of the wastes that cannot be treated for any other purpose is then converted into compost. However, the mining industry in Canada is usually not connected with large urban centers so that conversion of wastes of comparatively small mining towns must be considered. Because of the long distances from potential market centers and the small populations, no revenues from material recycling can be expected (scrap iron, glass). Energy production can be considered to reduce costs required by the process producing compost. Although the production of energy from wastes of small towns appears to be economically feasible, compost production from wastes of mining towns is preferable because of its use in abating environmental problems. In areas where compost production is not a primary requirement, energy production may be more feasible. Incineration is used in Prudhoe Bay, Alaska combining the required waste disposal and energy production used for heating the plant in winter and for the freezing of the plant floor in summer to prevent the frozen tundra subsoil from melting (5). This case, however, is essentially also an environmental problem.

A summary of the results of preliminary experiments done on converting municipal wastes of small mining towns into compost is given in this report. The principles of a process that can be considered in view of the Canadian climate are described.

SUMMARY OF RESULTS OF PRELIMINARY EXPERIMENTS ON THE COMPOSTING PROCESS

Apart from keeping costs for the composting process to a minimum, the solutions to two problems were the principal objectives of the experiments.

a. The solid municipal wastes have a high content of cellulosic matter (70%) and represent, therefore, almost a mono-typic material that is resistant to degradation. Because this material might not contain micro-organisms specifically degrading cellulose, the problem was to produce and maintain a microbial flora that would act specifically in this respect. The microbial population of sewage sludge is not as specifically cellulolytic as manure from herbivores so that initial experiments were done without considering the use of sewage sludge.

b. Although the initial part of the windrow composting process (degradation) could be carried out in an enclosed area and the process itself produces heat, it remained questionable if the windrow process could be continued during the low ambient temperatures of Canadian winters without problems.

The first experiments used wood waste (minus 5-mesh) in view of a possible supply of wood waste from saw mills that might be operating near the mining sites. These experiments (6,7) indicated that wood waste could primarily be used to improve the porosity of the garbage mixture. The type of aeration ("turning" or forced airflow) made no significant difference in the degradation of wood waste, and the rigid wood particle structure permitted the use of the maximum-moisture content that the waste could hold. A bacterial-concentrate addition to the waste affected the degradation, yet there is no clear evidence that it was an absolute necessity for the composting process.

The general point of view was taken, therefore, that the more heterogeneous the mix to be composted, the less the requirement for a continuous use of a bacterial-concentrate will be. Consequently, the microbial population might be maintained by recycling compost from the garbage itself.

Compost for recycling was obtained in another experiment (8). The results indicated that the process was sensitive to substances that might inhibit microbial growth. The compost that would be used for recycling was obtained from a mixture of shredded garbage, soil and a commercial compost. This combination had better results than the admixture of soil alone or that of a bacterial concentrate to the garbage.

The experiments indicated furthermore that forced-air aeration must be used for the degradation of garbage. The structure of paper matter constituting the principal component of the garbage is not as rigid and therefore less porous than that of wood waste.

The composting mass still has to be turned for the purpose of mixing the material not only to homogenize its composition during degradation, but also to expose all material to the high temperatures reached in a windrow.

The most preferable particle-size of the garbage was determined to be 1/2 to 5/8 inch (8). Also, it appeared feasible to evaluate the compostability of a mixture by subjecting in an incubator small volumes (4 L) of the mix to a temperature cycle simulating that of a composting windrow. This method was used to estimate the acceptable limits of garbage, sewage sludge, wood waste and recycled compost in an actual operation (9). The ratio of these components depended on the composition of the particular materials used in the experiment. The experiment showed that the garbage content could vary from 75% to 50% and the sewage sludge from 10% to 25%. A maximum of 20% wood waste could be used, while the recycled compost was kept constant at 10%.

The experiments had also indicated that the composting of high-cellulosic material might be affected significantly by temperature changes. The composition of the mixture that had appeared to degrade fastest was used in a lab experiment to estimate the effects of low ambient temperatures on degradation (10).

Field experiments to obtain such information would require that the winter temperature remained consistently low for the control period. Such experiments could be considered if the laboratory results were favourable. The experiment indicated that a significant part of a windrow would freeze at -20°C or lower temperatures. Furthermore, the results showed that heat had to be applied periodicaly during the process. In addition to low temperatures at the beginning, cooling occurs when a windrow is turned to mix the frozen and undegraded outside-layers into the warm core.

Apart from the requirement to supply heat, the process could be expected to proceed slower than at higher ambient temperatures. The solution to this problem was to store the mixture during the winter. This was only feasible if the bulkiness of the mixture could be reduced and foul odors that would be produced by garbage heaps during mild temperature spells be prevented.

Experiments were done on the degradation of briquettes of a hydraulically-compacted garbage mix which, according to the 'Kaspari process', would and did fulfill these requirements (11). The experiments indicated also that the briquettes not only tolerated but preferred a significantly higher sewage-sludge content (50%) than windrows. If only 20% sewage sludge would be available for mixing with the garbage, the degradation rate of the mixture could be increased significantly by adding a bacterial concentrate to the mix.

In order to design a process based on the results of the foregoing experiments and estimate the minimum costs, the Elliot Lake mining area was considered as an example.

ESTIMATE OF POTENTIAL COMPOST PRODUCTION OF ELLIOT LAKE, ONTARIO

The organic part of Canadian garbage consists of 70% paper matter, 10% garden wastes and 5% rags, plastics, etc., totalling 85%. The remaining 15% consists of 5% metals, 5% glass, ceramic materials, etc., and 5% ash and "dirt" according to v.d. Broek and Kirov (12). They estimated that the average garbage production is 2.2 lbs/cap/day.

The composition and production of garbage varies locally. For instance, Pavoni (13) arrives at a total of 62% organic matter (without plastics) and an average production of 5 lb/cap/day for U.S. garbage. The Ontario Ministry of the Environment estimates garbage production in the Toronto area at 4.5 lb/cap/day. In view of its distant location where the use of pre-packaged food can be expected to be above average, the 85% organic-matter content is assumed to be valid and a conservative estimate of 4 lb/cap/day garbage is used in the following:

The population of Elliot Lake is approximately 13,000 and is expected to double in the next decade. Consequently, the production of garbage can be expected to be, at least, 25 tons/day* increasing to 50 tons/day in the next decade. The average yearly garbage production increases, therefore, from 9,000to 18,000 tons. If 10,000 tons are considered, 1,500 tons would consist of non-compostables of which 500 tons would be scrap iron.

If a conservative reduction-factor, due to moisture loss, etc., of 2:1 is applied to the 8,500 tons of remaining organic matter, a present production of 4,250 tons of compost/year from garbage in the Elliot Lake area can be expected.

This production is increased to some extent by the inclusion of sewage sludge in the garbage mix. Fritz (14) provided data on the design capacity of waste-water treatment plants and the populations served in twenty areas. According to his data the average volume of waste water that can be expected to be treated per day per capita is 644 litres. Although this figure seems high, a population of 25,000 would produce only 5 tons of compostable solids per day according to the organic-solids content of waste water estimated by Gunn (15). The sewage factor is therefore neglected in the estimates. The water-to-solids ratio in septage is smaller than in waste water, but is very variable because it is dependent on many factors.

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* Note: 1 ton = 2000 lbs has been used throughout the report

PRELIMINARY DESIGN OF A PROCESS CONVERTING MUNICIPAL WASTES INTO COMPOST

A composting tractor, in conjunction with a tanker truck to distribute sewage sludge, could be used for composting garbage and sewage sludge during the summer in the field. However, the resulting particle size and extent of degradation remain questionable. Therefore, the use of a shredder system was selected because it could be expected to produce material having a particle size suitable for fast degradation.

The results of the experiments indicate only the principles for a waste-disposal compost-production process. The preliminary design of the process, described in this report, can serve as a basis for pilot-plant experiments.

The operating plant could be used as the means of total treatment and disposal of all municipal wastes. Septage or raw sewage-sludge could be dewatered using a centrifuge before mixing with the shredded garbage.

Due to the nature of the shredded garbage, a mixer particularly designed to mix sludge and garbage must be used.

The briquettes, stored during the winter and degraded to a certain extent, are shredded in a second mill. Because of the requirement to use approximately 1/2-inch particle size of shredded garbage and the difficulty to obtain this small size in one-stage milling, the action in this mill should facilitate mixing and particle size reduction.

The following optional machinery which would refine and improve the process are not considered in the tentative plan for the composting plant:

a) air classifier to remove the finest fraction for energy production (methane).

b) A ballistic- or gravity separator for the removal of glass, ceramics, nonferrous alloys, etc.

c) A trommel screen at the beginning of the process for the breakage of glass and other ceramic materials to save wear on the shredder.

d) A screen at the end of the process for the removal of plastics and other matter that has not degraded.

e) A dust removal system. Shredding of high-cellulosic material produces a considerable amount of dust which presents a health hazard.

f) A protective enclosure and vibration absorbance for the primary shredder. Garbage often contains small containers or spray cans that could explode in the primary shredder and the vibrations of the shredder could damage the building structure.

The overall process flowsheet is shown in Fig. 1. An arrangement of the principal equipment for the collection, shredding and mixing of garbage and sewage sludge or septage is illustrated in Fig. 2.

The 50-m wide front of the building has two truck-wide entrances. Garbage is dumped on the left on a 'receiving' platform. It is then pushed through an 8-m wide entrance onto a feed conveyor-belt using a small tractor equipped with a power-angle front-dozer blade (e.g. Kubota* Model B 6100 Diesel, four wheel drive). The garbage passes a levelling gate before it reaches a hand-picking station where non-compostables, hazardous to the shredder, are removed. These materials are sorted into hoppers. The contents of each hopper can be conveyed to a baler. Three hoppers are shown to separate different types of material, e.g., organics (rags, ropes, etc.) metal (wire, bars, etc.) and miscellaneous (containers with combustible matter, etc.).

The garbage is shredded in the (primary) Shredder-I (Williams P.C.P. Co., "down running", Model 340 K.) (16) and passes under a magnetic separator (Dings Co.) which removes the scrap iron. This material is also carried to the left on the waste conveyor-belt so that bales of scrap iron can be pressed and trucked out through the left side exit.

Similarly, the fraction collected by a ballistic-separator (B) is transported on this waste conveyor-belt and trucked out. The left entrance and exit are, therefore, for the removal of non-compostable material.

Sewage sludge or septage is collected in a tank and pumped to a centrifuge (Sharples, model P 3500) where it is de-watered. The shredded garbage is mixed with sewage sludge in a pugmill-type mixer (Sprout-Waldron or McLanahan Co.). The mixture is then carried to a secondary shredder (Eirich Co., Model TRC 22) where it is shredded to 1/2-inch particle size and mixed with briquettes that have been stored during the winter. These briquettes are pushed through wicket 'C' onto the same conveyor-belt that runs under the pugmill. The mixture produced by the secondary shredder is either discharged at 'D' to be windrowed during the summer, or it is carried to a baler (AMBACO, continuous hydraulic horizontal, Model 10445) to be compressed as briquettes during the winter. The direction of the conveyor-belt collecting the products from the secondary shredder can therefore be reversed.

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*Brand name of products are used in this report only as examples, but do not imply endorsement of the products or a selection based on experimental data.

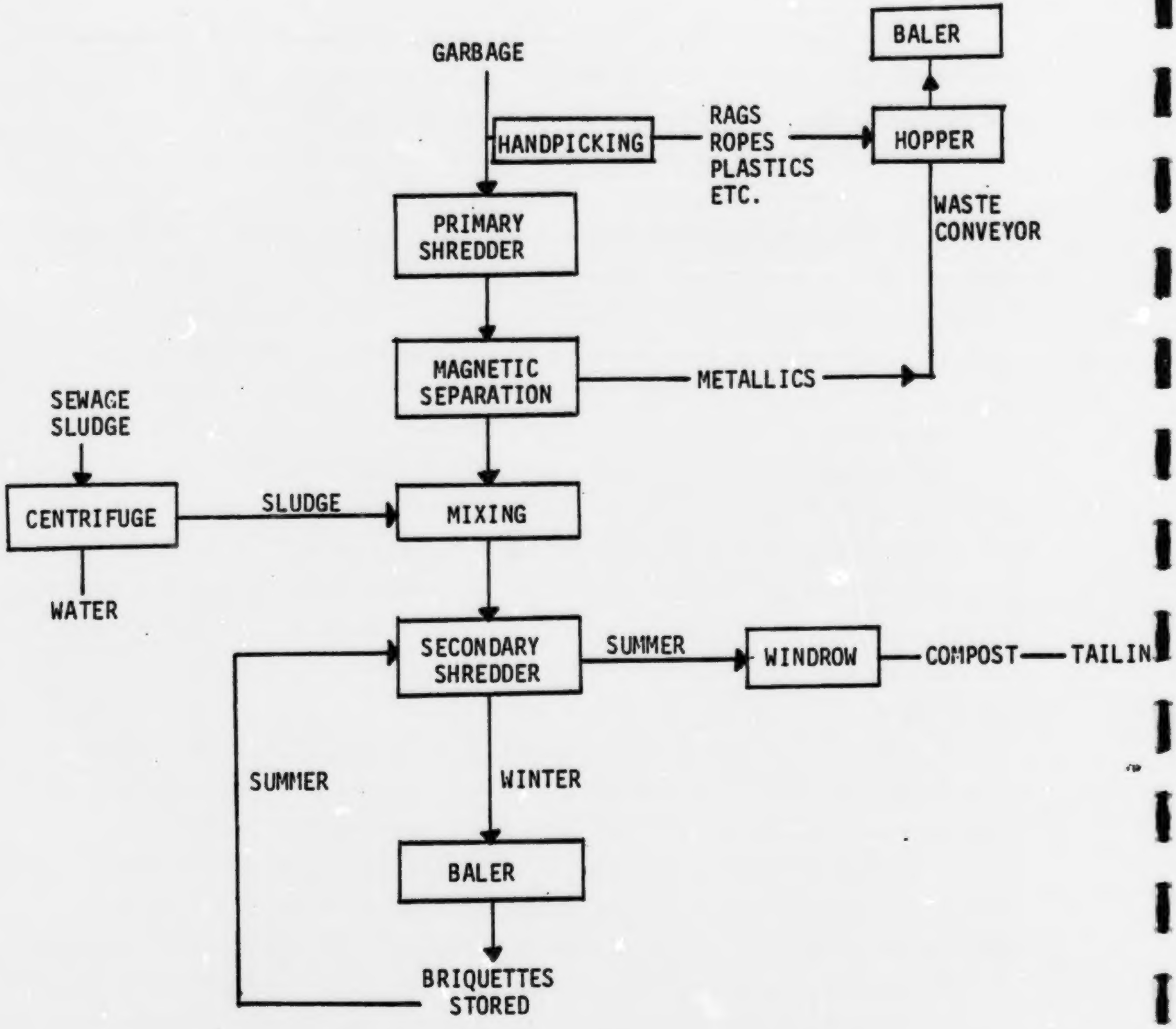
GARBAGE
HANDPICKING
RAGS
ROPES
PLASTICS
ETC.
BALER
HOPPER
WASTE
CONVEYOR
PRIMARY
SHREDDER
MAGNETIC
SEPARATION
METALLICS
SEWAGE
SLUDGE
CENTRIFUGE
SLUDGE
MIXING
WATER
SECONDARY
SHREDDER
SUMMER
WINDROW
COMPOST
TAILIN
SUMMER
WINTER
BALER
BRIQUETTES
STORED

Figure 1 - Schematic of composing

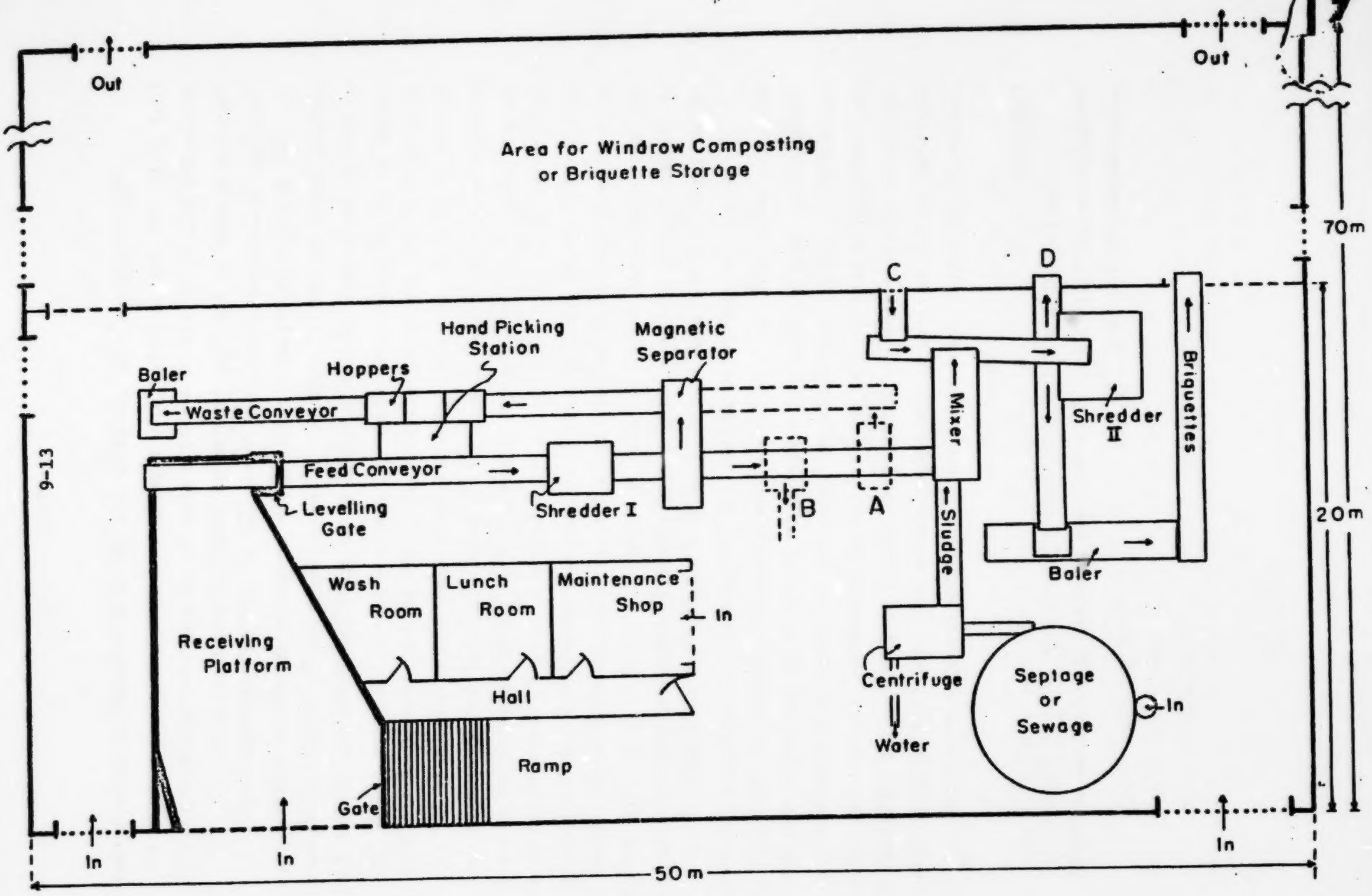
Out
Out
Area for Windrow Composting or Briquette Storage
70m
C
D
Hand Picking Station
Magnetic Separator
Baler
Hoppers
Waste Conveyor
Mixer
Shredder II
Briquettes
Feed Conveyor
Shredder I
Levelling Gate
B
A
Sludge
20m
Baler
Wash Room
Lunch Room
Maintenance Shop
In
Receiving Platform
Centrifuge
Septage or Sewage
In
Hall
Water
Ramp
Gate
In
In
In
50 m

Figure 2 - Tentative plan for a composting process plant

The briquettes from the baler are put on a slow-moving conveyor-belt from which they can be pushed onto a pallet. A fork-lift truck carries and stacks the pallets in the 50 m x 70 m area for storage during the winter. They are stacked in the same fashion as freshly-fired builing bricks. If required, collapsable steel-pole frames can be used as supports.

The size of the area required for briquette storage or windrow composting under shelter is estimated in Appendices 2 and 3. This size appeared to be similar to that used in Virginia in an operation producing compost from sewage by the windrow method (15). An overhead-travelling crane can do the major part of the moving and stacking while the tractor transports the briquettes to and from the area. The mix to be windrowed at discharge 'D' is transferred by a front-end loader into a truck with side-spreader which constructs windrows outside where a Beltsville-type forced-aeration windrow-composting system is used. It may be advantageous to have forced-aeration pipes in a part of the 50 m x 70 m area to produce a comparatively dry mix that can be used for mixing windrows outside in wet periods. In order to save space the "moving windrow" method designed by Spohn (17) could be used. The mixing of the material is achieved using a snow blower for transferring the composting material from one side to the other of a lane in the material.

After the completion of each run the lane moves sideways until gradually the whole mass has been mixed. The material processed per unit area is significantly greater than that occupied by conventional windrows. The method was designed originally to avoid the formation of a core in a windrow where the composting action is slow or could become anaerobic. However, the use of the forced-aeration pipes is recommended because it prevents the formation of anaerobic conditions even if the season may be wet. Moisture is controlled further by the admixture of the comparatively-dry shredded briquettes.

ECONOMICS

The following are the 1982 Costs of Municiple Garbage Disposal and Revegetation in Elliot Lake. This data was provided by N.K. Dave of CANMET Laboratories, Elliot Lake(18).

1. Town of Elliot Lake Municipal Garbage Disposal Costs:

Data -

Total garbage produced per year = 16,000 tonnes
Disposal yard lift = 5' (4'-6')
Dirt cover on top = 6" - 1'
Compact garbage density = 30 cu. ft/tonne

Capacity -

2 garbage collection trucks (residential) = 25 cu.yd. each
1 commercial truck for containers = 30 cu.yd. each
1 dump truck = 14 cu.yd. each

Costs - Collection -

Collection and haulage, 2 trucks, 8 h/day
5 days/week, 52 wk/yr @ $25.00/h each = $104,000.00

Commercial $30/h, 40 weeks = 48,000.00

Total: $152,000.00

Area covered - 5' lift $= \frac{16,000 \times 30}{5} = 96,000$ sq.ft.
= 2.2 acres
Cover fill required, 1' cover = 96,000 cu.ft. = 3,560 cu.yd.
= 254 loads
Cost of cover @ $30/load = $7,600.00
Cover spreading equipment, costs = $2,000/mth, 12 mth/yr. = $24,000.00

Total costs per year = $183,600.00
Total acres covered = 2.2 acres
Total cost/acre = $ 83,450.00

2. Revegetation Costs:

First Year -

@ 10-30 tonnes of lime/acre
40-50 lb seeds/acre = $1,000/acre

Next 2 to 5 years fertilizer application -

@ $200/acre per year = $800/acre

Total for five years: = $1,800-$2,000/acre

The costs for the composting operation(19) are estimated at $1.5 million capital and are estimated $341,250 for annual operating and maintenance. The operating costs are based on a requirement for 7 men on a 5-day week at $25/h.

DISCUSSION

Various ways and means have been considered to convert household garbage into compost at a minimum cost. A popular attempt is to solicit the cooperation of the relevant population for separating their garbage into an organic and inorganic fraction (two-bin system). Experiences with the willingness to cooperate and the actual cooperation level have been described by Heron (20). "Prior to implementing the program in the New York City case study area 20 per cent of the householders indicated a desire to participate. The average participation rate achieved on commencement was two per cent".

The use of a composter tractor and a tanker truck containing and dispersing sewage sludge on the garbage can be considered. The quality of the resulting compost, however, is questionable and the arrangement has to be tried out. This is feasible particularly to initiate a composting operation because the requirements do not involve the acquisition of permanent equipment. The use of low-grade or immature compost, however, will have an adverse effect on plant growth. Nitrogen, for instance, will be used by continued degradation of the compost and this will be detrimental to the plant growth (21,22). Furthermore, non-compostables are not removed, and the sorptive properties of compost that has a larger particle size than that produced by a shredder also remain questionable. If an operation of this kind is considered, it will probably have to be extended by screening the product.

The principal point to consider is that compost production from municipal wastes involves waste disposal even if it is confined to using the organic fraction collected in a two-bin system. It seems therefore logical to consider compost production as a means of waste disposal. Obviously, solid waste disposal using the land-fill method is most economical in remote mining areas, yet septage or sewage still has to be treated. In the process that has been described in this report, the costs for garbage and septage collection will still accrue. Non-compostables must be disposed of, and the costs for their disposal remain uncertain. However, the Eirich shredder crushes glass and other ceramic materials to a particle size less than 6 mm, so that this fraction of the non-compostable garbage may not require separate disposal. Scrap iron can be baled and stored for future economic use, so that costs for removal of the remaining non-compostables may be low. The principal maintenance costs for a landfill site can therefore be eliminated.

A reduction in costs for sewage or septage disposal may be more significant if a composting plant can serve this purpose. The total costs should in fact, be carried by both the municipality and the mining industry. To effect a true cost, the costs for current waste disposal should be deducted from compost process costs to arrive at the net cost. No information in this respect is available.

The proposed process is based on the production of 50 tons/day garbage for a projected population of 25,000. The present production can only be half that of the projected production. Minimum requirements must, therefore, be considered to begin a composting operation, particularly because the existing waste-disposal system has to be changed gradually.

Since the size of the building cannot be reduced, the 50 x 70 m area would be sufficient for windrow composting for at least, 25 days (see Appendix 3). During the first year a continued windrow-composting process could, therefore, be attempted notwithstanding the problems that are involved. Heating of the forced-air aeration system and the required increase of labour would not be permanent. Heat has to be supplied to begin the process and more than a normal amount of mixing will be required to produce a homogenous material. However, although investment costs for a secondary shredder, baler and possibly a centrifuge may be saved initially this way, it would not reduce the operating costs.

The costs of the process may appear to be close to that using silos which would require a smaller building. Apart from the difference in costs which were estimated to require, at least, a 2-million dollar investment (23), the outer layers of the composting mass can be expected to freeze and stick to the walls of the silos. It is possible that this will not impede a continuous composting process in any way so that the feasibility of adopting this type of process can only be evaluated by actual scale experiments conducted by a silo-producing company visualizing that such a process will be less costly and will produce less problems during the winter than that which stores the wastes as hydraulically-compacted briquettes.

The experiments that were carried out, were concerned with the degradation of high cellulosic wastes (82% paper matter) particularly during low ambient temperatures. Problems of trace metal content or pathogen destruction were not included in the study.

CONCLUSIONS

The work reported in this report has shown the following:

- Various methods have been demonstrated to convert municipal garbage into compost.
- Compost production could be considered as means of waste disposal.
- Maintenance costs for landfill sites could be eliminated.
- Scrap metal could be stored for future use.
- Reduction in costs for sewage and septage disposal could be significant if combined with composting.
- Total costs should be borne by both municipality and the mining operation(s).
- A managed site is produced which would be environmentally acceptable to the mining community.

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APPENDIX 1

ESTIMATE OF DENSITIES OF COMPACTED GARBAGE

The densities of the garbage mix at various stages of compaction are compared in the following.

The density of a briquette pressed at 20 kg/cm^2 is 475 g/L (dry wt).

The density obtained by manually compressing CANMET garbage (non-compostables removed) in a polyethylene tank was 23 kg/50 gallons (189.25 L). Assuming 15% moisture the dry weight of this garbage was (0.85 x 23 kg =) 19.55 kg. The density was, therefore, 103.3 g/L.

The density of garbage used for estimating the space required for windrow composting (see Appendix 2) was 200 lbs/cu yd or 90.7 kg/764.6 L (0.85 x 90.7 kg = 77.10 kg dry wt).

The density was thus (77,100/764.6 g/L =) 100.8 g/L. This value includes 15% metals, glass, etc., which may have an average S.G. 4 (g/cm^3). This occupies therefore a negligible space so that the actual density is approximately 85.7 g/L.

The compaction ratio produced by the baler is 475/85.7. or approximately 5.5:1. This ratio increases to approximately 7 concerning freshly shredded garbage that has not been compacted except by its own weight.

APPENDIX 2

ESTIMATE OF SPACE REQUIRED FOR STORING BRIQUETTES

Treatment of 50 tons/day garbage results in 42.5 tons/day organic matter (pg. 6). According to experimental results described in a previous report (7), the assumption is made that the general moisture content of this garbage is 15%. The production of organics that can be converted into compost is thus 36.125 tons/day (dry wt).

The number of months that garbage will have to be stored is assumed to be six. The total dry weight of garbage to be stored is therefore $6 \times 30 \times 36.125$ ton = 6502.5 tons = approx. 6×10^6 kg.

The density of a dry briquette was determined to be 475 g/L. The size of the briquettes can be expected to be 26 in. x 18 in. x 36 in. (high) according to the bale sizes produced by the AMBACO, model 10445, continuous hydraulic horizontal baler. The volume of one briquette is then approximately 276 L. Its dry weight is 131 kg. The total amount of garbage produced in 6 months must therefore be stored as $\frac{6\text{x}10^6}{131}$ = 45800 briquettes or $\frac{45800}{180}$ = approximately 254 briquettes per day. The AMBACO baler compacts 25 cubic feet (0.7 m^3) per ram stroke according to the information from the manufacturer. Because one briquette has a volume of approximately 270 L (0.276 m^3), one ram stroke compacts the material to be cut into 2.5 briquettes. The machine can provide apparently a maximum of 4 strokes per minute so that 254 briquettes can be pressed in a day-shift without difficulty.

For storage space of a 26 in. x 18 in. x 36 in. (high) or 66 cm x 46 cm x 92 cm briquette, a floor space of 50 x 70 cm and height of 100 cm is assumed. The horizontal rows are separated by the thickness of the pallets which carry several briquettes. Space along the length and width of the briquette will provide air circulation. If a 3-m wide lane is left open along the front of the storage area, its length is 67 m. This would allow storage of 96 briquettes. Stacking them 6-m high allows the storage of 576 briquettes. A total of 45800 briquettes have to be stored which will occupy therefore $\frac{45800}{576}$ = 80 rows parallel to the side, or a width of about 40 m. This leaves an open lane of 10 m on the left of the storage area. Obviously, the briquettes could be stacked in such a way that 5-m wide lanes are left open along the sides of the building so that the oldest briquettes are available for shredding at the beginning of the windrowing season, or four 2.5-m wide lanes could be constructed.

APPENDIX 3

ESTIMATE OF SPACE REQUIRED FOR THE WINDROW COMPOSTING PROCESS

Since the conventional windrow size is uneconomical, the "moving-windrow" method designed by Spohn (17) should be adopted. This method requires two lanes of approximately 2-m wide. One lane is used by a snow-blower transferring (and thus mixing) the material from one side of the lane to the other. The second lane is used by a truck equipped with a side-spreader which deposits the fresh material. Because the slopes of the windrow are approximately 2:1 and its height is 150 cm, 75 cm of unoccupied space is added to the 2 m occupied by the lanes. Leaving the same 3-m lane in front of the 50 x 70 m area as for briquette storage, a floor space of (50-2.75)m x (70-3)m = 3166 m^2 can be occupied by the moving windrow. This will occupy a volume of 3166 x 1.5 m^3 = 4750 m^3.

The bulk density of raw garbage varies greatly and depends on compaction. An average value of 200 lbs/cu yd will be used in the estimate because Kirov (12) estimates the density at 215 lb/cu yd and Pavoni at 175 lb/cu yd (13). Accordingly 50 tons/day would occupy a volume of approximately 500 cu yd or 380 m^3. The total area could therefore contain the product of $\frac{4750}{380}$ = 12.5 days or nearly two weeks of 50 tons/day production. Presently, garbage production of approximately twice as many days can be treated because half the quantity of garbage is produced. Furthermore, garbage reduces in volume considerably during the composting process so that more than the estimated production could be windrowed. It is not known if the space is sufficient for windrowing at present, because of lack of information on the degradation rate during the winter.

However, in an actual operation the area will gradually be filled with briquettes during the winter which will be depleted during the summer. Although the output will contain twice as much garbage (50 tons/day fresh garbage and briquettes produced from 50 tons/day garbage), the actual volume of the shredded briquettes added to the fresh garbage mixture is not known.

10-1

BRIEF DESCRIPTION OF THE MADAWASKA PROCESS

MADAWASKA MINES LIMITED

MILL PROCESS

The process consists of grinding, dewatering, sulphuric acid leach, acid filtration, tailings neutralization, clarification, ion exchange, precipitation and drying and packaging of the product.

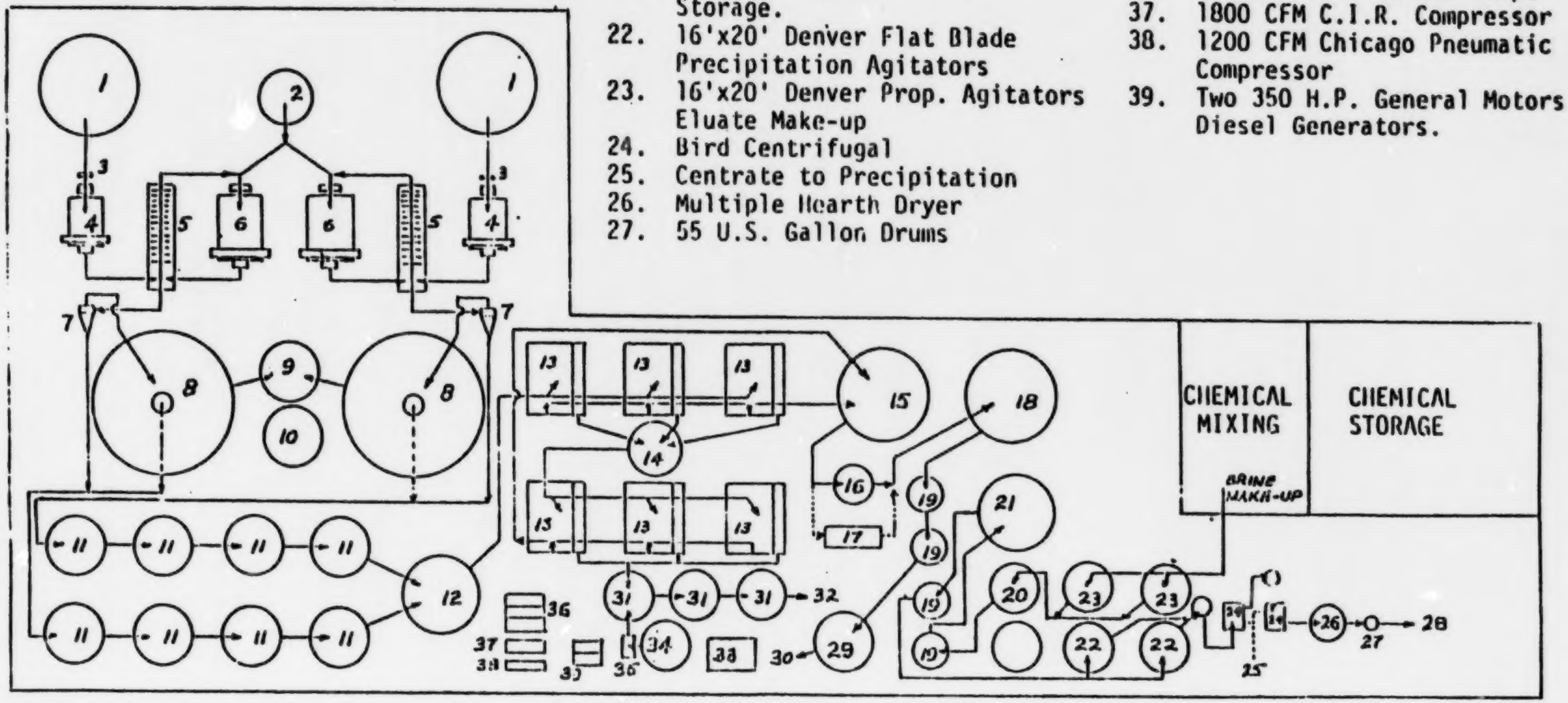
Grinding	- two circuits with one rod and pebble mill - ore size is ½ inch - pebble size is between 1 3/4-inch and 3 inches
Dewatering	- two circuits with one 15-inch cyclone and 45-foot thickener
Leaching	- is done at 65% solids at a pH of 1.8
Acid Filtration	- primary and secondary filtration removes uranium bearing liquor from the leach residue
Tailings Neutralization	- leach residue is neutralized with lime to pH 8.5 prior to discharge to tailings dam.
Tailings Disposal & Treatment	- liquid effluent decanted from the tailings pond runs into the mill for treatment. Prior to treatment with barium chloride a portion of the effluent is reused in the milling process. The barium radium coprecipitate is permanently captured in a settling tank
Clarification	- both a Himsley sand clarifier and an Eimco precoat clarifier are used to remove suspended solids from the pregnant leach liquors
Ion Exchange	- four Permutit stationary bed ion exchange columns are used either on two or three column absorption. Both Rhom & Haas IRA 400C and Dowex 21K ion exchange resins are used.
Precipitation	- is done by batch process using magnesia to raise pH to 7.
Drying	- precipitate is pumped through a two stage centrifuge system to produce a yellow cake slurry approximately 55% solids. This is dried in a multiple hearth dryer.
Product	- Yellowcake, approximately 80% U_3O_8, shipped in 55 U.S. gallon metal drums. Maximum drum weight is 1,000 lbs.

** A flow sheet showing the mill process is attached.

MADAWASKA MINES LIMITED

1500 TPD Acid Leach Plant
General Arrangement & Flow

1. 36'x40' Fine Ore Bins
2. 12'x16' Pebble Bins
3. Ramsay Weightometers
4. 6'x8' Hardinge Rod Mills
5. 5'x26' Dorr Classifiers
6. 9'x11' Hardinge Pebble Mills
7. 15" Krebs Dewatering Cyclone
8. 45' Dorr Thickeners
9. 16'x22' Grinding Solution Storage
10. 17'x20' IX Backwash Water Storage
11. 18'x45' Pachucas
12. 24'x20 Denver Surge Agitator
13. 11'-6"x16' E Long Filters
14. 10'x8' Denver Agitator
15. 30'x30' Filtrate Storage
16. Himsley Pressure Clarifier
17. 8'x16' Eimco Drum Clarifier
18. 24'x25' Pregnant Surge Tank
19. 9'x16' Ion Exchange Columns
20. 16'x20' Fresh Eluate Storage
21. 24'x22' Recycle Eluate Storage.
22. 16'x20' Denver Flat Blade Precipitation Agitators
23. 16'x20' Denver Prop. Agitators Eluate Make-up
24. Bird Centrifugal
25. Centrate to Precipitation
26. Multiple Hearth Dryer
27. 55 U.S. Gallon Drums
28. To Refinery
29. 20'x16' Barren Effluent Surge Tank.
30. Barren to Re-use Or Waste
31. 10'x12' Denver Prop. Agitators Tails Neutralization
32. To Fill Plant &/or Disposal
33. 30 Ton Lime Hopper
34. 75 Ton Lime Storage
35. Omega Lime Slaker
36. 31"x13" C.P. Vacuum Pumps
37. 1800 CFM C.I.R. Compressor
38. 1200 CFM Chicago Pneumatic Compressor
39. Two 350 H.P. General Motors Diesel Generators.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CHEMICAL MIXING
CHEMICAL STORAGE
BRINE MAKE-UP
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POOR

QUALITY

ORIGINAL